SNC

1, Complexe Desjardins
P.O. Box 10
Desjardins Postal Station
Montréal, Canada H5B 1C8

Telephone (514) 282-9551

Cable Address SNCINC
Telex 055-60042

SNC Inc.

January 27, 1984

SNC File No.: 6733

Dr. David Fung
Department of Energy, Mines and Resources
CANMET, Energy Research Laboratories
555 Booth St.
Ottawa, Ontario
K1A 0G1

Re: Application of Fluid-Bed Gasification Technology and Absorbents (Dolmite, Limestone, Oil Shale) for Utilization of High Sulfur New Brunswick Coal
Revision of the Final Report

Dear Dr. Fung:

We are enclosing seven (7) additional copies of the final report, previously approved by you.

With kind regards.

Yours truly,

Zafar A. Siddiqi
Project Manager

ZAS/gd
Encls.

cc: Ms. Gaetane Dagenais
Department of Supply and Services
Science Centre
11C1 Place du Portage, Phase III
11 Laurier Street
Hull, Quebec
K1A 0S5

COMMENTS RECEIVED FROM CANMET
ON THE REPORT

1. The analysis of sulfur forms shown in Table 4-3 differ from the values reported in Table 6-2.

Should be:	pyritic sulfur	52.88%
	sulfate sulfur	11.86
	organic	34.47
	$S^{=}$	0.80
		100.00%

2. Indicate the bed height of various experimental runs.

3. Results in Appendix B indicate CH_4 value to be about 2%. This indicates that the gasifier is probably operating halfway between fixed-bed and fluid-bed mode. Please comment and add appropriate explanation.

4. Figure 6-1 indicates that there is a large variation in H_2 concentration particularly for Runs #830705, 830707, 830714. Is there an explanation for such a behaviour? If so, kindly include or at least comment on such a large variation.

5. Complete the analysis/results in Appendix B. Values in some of the runs are missing.

6. Please complete proximate analysis given in Appendix B, as done for Run #830606 (p. 89).

7. There is no oxygen in the synthesis gas. Is there no oxygen at all or it is not measured?

8. In addition to the results presented, the data must be expressed for each run as shown in the attached table.

9. P. 12 shows steam-to-air ratio for various runs. Air/coal (maf) ratio which is most important is not shown. Also next important steam/coal (maf) is missing.

10. P. 64: reason for the CO assumption.

11. Page 70: include conclusion #2 "To counteract..... coal particles", in the following recommendation section.

12. In Conclusion, Section 7, comment on

    - effect of air/coal ratio
    - effect of steam/coal ratio
    - effect of inert
    - effect of physical characteristic properties of inert

13. P. 71: Line #9"Modifications to the process are required for gasification of Minto coal" may be deleted.

RESPONSE TO COMMENTS

1. The values given in Table 6.2 are correct. These values are based on the original analysis from Analyst A on an as received basis. Table 4.3 has been modified and the sulfur forms in Minto coal are reported on an as received basis. In the earlier submission the analysis was corrected to a dry basis with values rounded off to two decimal places as in the original analysis.

2. The bed height was regulated by solids overflow through the overflow port (module 4, p. 16) which is located 1.2 m above the distributor plate (p. 42). The settled bed depth is 0.6 to 0.7 m.

3. The gasifier was operated at fluidization velocities well above the minimum. However, agglomerate formation had an adverse effect on the quality of fluidization and mixing in the bed. CO_2 from the calcination of limestone has a diluting effect on the CH_4 concentration (Fig. 6.1).

4. The fluctuation in the H_2 concentration in the synthesis gas is referred to on p. 60. One can only speculate on an explanation for this behavior. Agglomerate formation and the periodic development of local zones of high temperature could increase the H_2 output from the H_2O - C reaction.

5. We have included all the gas analyses which we have in Appendix B.

6. The proximate analyses in Appendix C now include values for total sulfur.

7. Oxygen was present in the synthesis gas in trace amounts. Quantitative measurement was not possible using our gas analysis system. Sulfur dioxide concentrations were not monitored in the synthesis gas. 'For the reducing atmosphere of partial oxidation, the sulfur in the gaseous phase appears principally as H_2S whereas in the oxidizing atmospheres of complete oxidation the sulfur appears principally as SO_2. More appropriately it is a matter of the H_2S - S - SO_2 equilibrium, though other compounds also make an appearance. As a rule of thumb, however, H_2S or SO_2 do not coexist; rather it is a matter of H_2S - S or else S - SO_2.' From Hoffman, E.J., 'Coal Conversion', p. 173, Modern Printing Co., Laramie, WY, 1978.

8. As requested, the results are summarized in tabular form in

Appendix D.

9. The ratios given in Table 3.1 are set values. An explanatory note on the operating strategy is given below Table 3.1. The bed temperature was controlled by adjusting the coal feed rate. Therefore, the air/coal and steam/coal ratios were not set values. Operating values for these ratios are given in Appendix D.

10. Conversely, there is no reason to assume that all of the O in the CO came from the air. The CO assumption was part of one calculation scheme which was used to estimate the flowrate of the synthesis gas. In the estimation of the sulfur capture three calculation schemes were used in an attempt to block in the estimates. Taken together, the three schemes made use of the analyses of all the incoming and outgoing streams. In carrying out the material balances, the most questionable assumption made was steady-state operation of the gasifier. If steady-state operation had been achieved, the validity of the CO assumption could have been checked by comparing the estimates of the sulfur capture obtained by the three calculation schemes.

11. This modification has been made.

12. The work concentrated on the problem of particle agglomeration which precluded the attainment of steady operating conditions in the fluidized bed. Since steady state conditions were not attained, the material balance results can only be interpreted qualitatively and no firm conclusions should be drawn. Some of the material balance results are plotted in Appendix D to indicate trends.

13. This statement stands. The work has demonstrated that modifications are required to overcome the problem of particle agglomeration.

APPLICATION OF FLUID-BED GASIFICATION TECHNOLOGY AND ABSORBENTS (DOLOMITE, LIMESTONE, OIL SHALE) FOR UTILIZATION OF HIGH SULFUR NEW BRUNSWICK COAL

Project Manager: Z.A. Siddiqi
Principal Investigators: H.A. Becker and R.K. Code
Co-investigators: R.W. McCleave and J.R. Stephenson

A report by

Technical Development Department
The SNC Group
1, Complexe Desjardins
Montreal, H5B 1C8

in association with

Department of Chemical Engineering
Queen's University at Kingston
Kingston, K7L 3N6

Work on this project was conducted under the auspices of Energy Research Laboratories, CANMET, Energy Mines and Resources Canada

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ABSTRACT

Gasification tests on a high-sulfur, caking coal from the Minto (New Brunswick) field were carried out in a pilot-scale atmospheric fluidized bed reactor with a square cross-section of 0.155 m^2 and a height of 3.3 m. Havelock (New Brunswick) limestone was used as a sorbent for sulfur capture. The range of the Ca/S ratio in the combined coal and limestone feed was 0 to 3 mol/mol. Steam and air were injected into the bed at steam-to-air ratios of 0.087, 0.22, and 0.38 mol/mol. The superficial fluidization velocities were 1.3, 1.6, and 2.3 m/s at the bed temperature of 900 °C. Although agglomerate formation in the bed resulted in short running times and prevented the attainment of steady operating conditions, approximate sulfur balances were carried out to obtain an indication of the redistribution of the sulfur in the coal among various sulfur-containing species in the product streams. The major species conversion in the gasification process was from pyritic sulfur in the coal to $S^=$ in the bed solids and cyclone dust. The H_2S concentration in the synthesis gas is reduced significantly when limestone is used as a sorbent. At a Ca/S ratio of 3 the sulfur capture is of the order of 90 %. Process modifications involving pretreatment of the coal and changes in the configuration of the bed and mode of coal dispersion are outlined in suggestions for further study on the problem of agglomerate formation in the gasification of Minto coal.

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1. INTRODUCTION

1.1 Scope of work

The work described in this report was undertaken to establish the suitability of utilizing New Brunswick coals (high sulfur, caking coals) for the production of low calorific gas in a fluid-bed gasifier. Since these coals have a high sulfur content, the work statement called for an examination of dolomite, limestone and oil shale as potential absorbents for in-situ desulfurization. In order to undertake this work the existing pilot plant at Queen's University was modified for operation as a gasifier. The pilot plant and support facilities are described in a recent technical report (Becker and Code 1982).

1.2 Specific tasks and objectives

The assignments outlined in the work statement of the contract are reproduced below.

1. Conduct a literature survey between the period of 1965 to 1982 on application of fluid-bed gasification technology for high sulfur coal using oil shale as absorbent.

2. Perform fluidized gasification studies with Minto coal (or its equivalent) and sulfur absorbents in a 35-cm diameter reactor having a capacity of about 50 kg per hour so that the following

information may be obtained:

- o composition of the product gas
- o carbon and sulfur contents in the product gas
- o carbon and sulfur contents in the carryover dust (cyclone catch)
- o carbon and sulfur contents in the bed material
- o sulfur content in the condensate
- o coal feed rate
- o fluidizing air flow rate
- o steam feed rate
- o absorbent feed rate, if used on a continuous basis.

3. Procurement and preparation of raw material which may include crushing, grinding and screening of coal, limestone, dolomite and oil shale.

4. Checking and calibration of various instruments, such as feeders, flowmeters, thermocouples, pressure gauges and other safety devices.

5. Perform preliminary runs to check the operability of the 35-cm gasifier system and make changes as necessary using the following experimental parameters:

Bed temperature	800 - 900 °C
Air/steam/carbon ratio	2-4: 0.2-1: 1
Fluidizing velocity	2 - 4 times minimum fluidizing velocity
Coal particle size	4 - 0 mm (with or without fines)
Bed depth	0.5 - 1.5 m
Absorbents	1.5 - 3 times stoichiometric requirements

Analyze the products, i.e., gas, cyclone catch, and bed material so that mass and energy balances can be made.

6. Perform 20 short runs (2-5 hours duration) to ascertain the important operating parameters such as coal feed rate, air/carbon ratio, steam/carbon ratio, bed temperature, etc. and the absorbent performance.

7. Perform 10 runs (8-12 hours duration) to study the effect of important operating parameters from Task 6 in detail on the gas quality and gasifier performance. These runs should use the best absorbent material as determined from Task 6.

8. Analyze the data and outline important recommendations for future work and scale-up.

2. LITERATURE REVIEW

2.1 Background

In general, the problem associated with sulfur impurities has been approached either by removing sulfur prior to gasification (beneficiation or washing techniques) or employing a tail-end purification method using a suitable solvent or sorbent. These conventional 'add-on' purification methods although effective, are inherently energy-inefficient and are also expensive (Hoffman and Truett 1972; Adams et al. 1977).

To alleviate these problems, several high-temperature desulfurization or hot gas clean-up systems have been proposed (Kohl and Riesenfield, 1980). A further refinement has been the 'in-situ' technique where gasification and desulfurization are carried out simultaneously in the gasifier by use of suitable sorbents.

A literature review of existing or developing processes which employ such 'in-situ' methods, pertaining to the processing of high-sulfur caking coal in a fluidized-bed gasifier configuration, is presented below.

2.2 In-situ Fluidized Bed Gasification/Desulfurization

Since desulfurization is an integral part of the processing system, sorbent selection is of relatively greater

importance compared to conventional 'add-on' methods. Relevant considerations are its thermodynamic stability at the gasifier operating conditions, resistance to attrition, acceptable emission properties, regenerative characteristics, and economic availability (O'Niell and Keairns 1977). These requirements tend to restrict the choice of sorbents for in-situ gasification/ desulfurization methods application.

The most widely used sorbents are limestone, dolomite and other calcium-based materials. Others such as FeO/silica have been used with moderate success (Oldaker et al. 1975) but in an 'add-on' purification scheme. The potential of oil-shale as a sorbent has also been suggested (Schacter 1979; and Wilson et al. 1980), however, its use has been restricted to combustion mode only. In preliminary experimental runs, Schacter's results indicated that the inorganic component of the oil-shale comprising $CaCO_3$, dolomite, silica, etc., is responsible for capturing sulfur in solid form during combustion. Sulfur is removed with the inorganic ash components. A detailed literature search on the utilization of oil-shale as a sorbent in a fluidized-bed gasification mode, revealed that information on this area is scanty, virtually non-existent.

Limestone and dolomite have been tested in some of the well-known fluidized-bed gasification processes, for example CO_2-Acceptor (Curran 1967, 1975), and Westinghouse (Archer et al. 1973; Holmgren and Salvador 1975).

In the CO_2-Acceptor process, calcined limestone CaO or dolomite CaO·MgO is used which not only captures the sulfur but also provides heat for the endothermic reactions. About 76% of the heat required in the gasifier is generated when CO_2 (one of the gasification products) combines with MgO·CaO to form MgO·$CaCO_3$ (Fink 1973). Sulfur retention has been claimed to be up to 70%.

In terms of coal feed, the CO_2-Acceptor process has been tested on low-rank lignite or sub-bituminous coals such as North Dakota lignite, Montana and Wyoming coals. These coals are essentially non-caking. If caking coal has to be used, a pretreatment step is necessary which normally reduces the thermal efficiency of the process by 6-8% (Wen and Desai, 1979). Additionally, the operating conditions (800-830 °C, 10-20 atm) give low gasification rates for bituminous coal. Higher temperatures to accelerate the reaction have resulted in melt and regenerative problems in the dolomite sorbent system (Henry and Louks 1971, Curran 1975).

Westinghouse, in a two-stage gasifier system, aimed primarily for producing low Btu gas for combined cycle power generation, has tested lime/dolomite as a sorbent.

Crushed coal after drying is introduced in a 'recirculating' bed devolatilizer/desulfurizer operative at 700-925 °C where lime sorbent and/or char is circulated in the

central draft tube of the vessel. These solids circulating at typically 100 times the coal feed rate effectively prevent or control agglomeration as the caking coal devolatilizes and passes through a sticky phase (Archer et al. 1978). Hot devolatilized char is continuously transferred to the second vessel where it is gasified at about 1050 °C.

Westinghouse claims to have gasified highly caking Pittsburgh coal without any pretreatment, but their devolatilizer is essentially a pretreater operating at high temperature; though their novel recirculating bed has a positive effect in controlling the agglomeration tendency of caking coals (Probstein and Hicks 1982).

Recently, emphasis has been directed towards the one-stage Westinghouse process (Holmgren et al. 1981) where devolatilization, gasification, ash agglomeration and separation are carried out in a single vessel. Recent tests have demonstrated that in this 'recirculating' type fluidized-bed gasifier, moderately caking coals can be gasified without any pretreatment (Meyers 1981).

The Institute of Gas Technology have gasified caking bituminous Kentucky coal in their recirculating type U-Gas gasifier (Myers 1981).

Recently, a Western Canadian caking coal of swelling

index 7 was gasified in a turbulent fluidized bed using sand as a heat and dispersion medium (Guitterrez and Watkinson 1982). The gasifier was operated at superficial velocities of 1-1.4 m/sec, where the bed was violently agitated or slugged. Also, more than normal N_2 had to be injected with the feed coal to prevent agglomeration.

It is evident that processing of a caking coal in a conventional fluidized bed has frequently been beset by difficulties caused by uncontrolled formation of agglomerates. The method most commonly used or proposed is to subject the coal to preoxidative pretreatment (Forney et al. 1967). This method, as mentioned earlier, is energy-inefficient

Alternatively, the key to feeding the caking coal directly into the fluidized bed is in achieving fast dispersal of raw coal particles in the dry char that comprises the bed (Yerushalmi 1977). These may be achieved through high velocity injection, or by creating high turbulence in the vicinity of the coal inlet port. These concepts are the subject of an active development program for fluidized-bed gasification (Chowdhury 1983).

3. THE PILOT PLANT

3.1 Preparatory work

Experience gained in earlier AFBC operation of the pilot plant indicated that the range of operating conditions specified in the work statement could be met. However, a number of modifications were required for operation of the pilot plant as a gasifier. Major modifications completed include:

- a new distributor plate and windbox for the dispersion of air and steam
- a quench cooler at the exit from the freeboard of the gasifier
- an incinerator for burning the synthesis gas prior to discharge to the stack
- a line to supply steam to the windbox.

Gas sample ports and probes were installed in the exit duct from the cyclone and in the discharge duct from the synthesis gas incinerator. Instrumentation for combustion gas analysis was in place. The gas chromatograph was moved from the analytical laboratory and installed in the pilot laboratory for analysis of the synthesis gas.

Preparations were made in our analytical laboratory for the proximate analysis and sulfur analysis of the coal feed and samples of the cyclone product and bed material. Arrangements were made with a commercial laboratory for analyses for carbon, hydrogen, nitrogen and specific sulfur-containing species in selected samples of the solids.

Supplies of Minto coal (- 3/16 in) and Havelock limestone (-16 +100 mesh) were prepared as feed materials. Arrangements were made with the St. Mary's Cement Company to obtain a supply of oil shale from their quarry at Bowmanville.

3.2 Limitations on the scope of the tests

Particle agglomeration in the bed severely restricted the scope of the tests. The accumulation of agglomerates reduced the quality of fluidization and resulted in a loss of temperature control in regions of the bed where the mixing was poor. This resulted in short running times under gasification conditions and precluded an approach to steady operating conditions. The problem also increased the turnaround time significantly as it was necessary to remove the solids from the bed and drop the windbox in order to clean agglomerates from the wall of the reaction chamber and the distributor plate. Startup from cold conditions increased the time on the front end of the next run.

A number of unsuccessful attempts were made to overcome this problem. The use of a dry coal feed and different limestone-to-coal ratios had no significant effect on the rate of agglomeration. The feed mode was changed from underbed to overbed in an attempt to improve the dispersion of the coal in the bed. The problems presented by particle agglomeration are discussed in more detail in later sections of the report.

Tabular summaries of the conditions of the tests and the number of samples collected from the product streams are presented in Table 3.1 and Table 3.2. Operating difficulties due to agglomerate formation were indicated by a rapid disparity between the thermocouple readings at different levels in the bed. The experimental effort concentrated on this problem and only Havelock limestone was used as a sorbent in the tests.

3.3 The gasifier

A flow diagram of the pilot plant, modified for operation as a gasifier, is illustrated in Fig. 3.1. The presentation is limited to a brief review of important features of the system since a detailed description of the pilot plant is given in a recent technical report (Becker and Code 1982). Modifications for operation of the pilot plant as a gasifier are described in more detail.

The main unit in the system is the fluidized bed reactor (gasifier or combustor), Fig. 3.2, which consists of ten modules. The chamber cross-section is nearly square, 381 mm by 406 mm, 0.155 m^2 cross-sectional area, in all modules except the top module which is transitional. From bottom to top, the modules are:

1. The windbox, a new redesigned unit for the injection of air and steam. A description of the windbox is given in a later subsection on modifications.

Table 3.1. Set conditions for operation of fluid-bed gasifier

Run	Superficial Fluidization Velocity at 900 °C, m/s	Steam-to-Air Ratio, mol/mol	Calcium-to-Sulfur Ratio, Ca/S
830601	(2.3)	-	3.0
830602	(2.3)	-	3.0
830606	(2.3)	0.087	3.0
830608	(2.3)	0.087	3.0
830610	(2.3)	0.087	3.0
830613	1.3	0.22	3.0
830614	1.3	0.22	1.5
830616	1.3	0.22	1.5
830620	1.3	-	1.5
830621	1.3	0.35	1.5
830623	1.6	0.38	1.5
830627	1.6	0.38	3.0
830705	1.6	0.38	3.0
830707	1.6	0.38	1.5
830712	1.6	0.087	3.0
830714	1.6	0.38	0.0

For runs 830601 to 830627 the gasifier was operated in the underbed feed mode. For runs 830705 to 830714 the gasifier was operated in the overbed feed mode.

The fluidization velocity was in the range of 2.3 m/s for runs 830601 to 830610. The coal and limestone feed rates were fixed and the bed temperature was controlled by adjusting the air flow rate. The steam flow rate was concurrently adjusted to maintain the steam-to-air ratio at the set value. For runs 830613 to 830714 the fluidization velocities and the steam-to-air ratios were maintained at the set values and the bed temperature was controlled by adjusting the coal feed rate. The limestone feed rate (runs 830613 to 830712) was adjusted to maintain the Ca/S ratio at the set value.

Table 3.2. Summary of number of samples collected from exit streams.

Run	Operation at Gasification Conditions,hr	Indication of Agglomerate Formation,hr	Number of Samples			
			Ash	Cyc.	Syn. Gas	Flue Gas
830601	0	-	0	0	0	0
830602	0	-	0	0	0	0
830606	2	.	3	6	3	0
830608	2	0.7	7	8	4	0
830610	0.5	.	1	1	0	0
830613	0.7	0.7	1	1	0	0
830614	2	1.3	4	2	3	0
830616	1	0.3	0	0	0	0
830620	0	-	0	0	0	0
830621	2	1	3	1	3	1
830623	2	2	4	3	4	1
830627	3	2	6	4	4	0
830705	5	1.5	10	5	11	2
830707	2	1.3	3	2	6	0
830712	2	2	5	3	5	0
830714	1.5	1.5	4	1	6	3
	Total		51	37	49	7

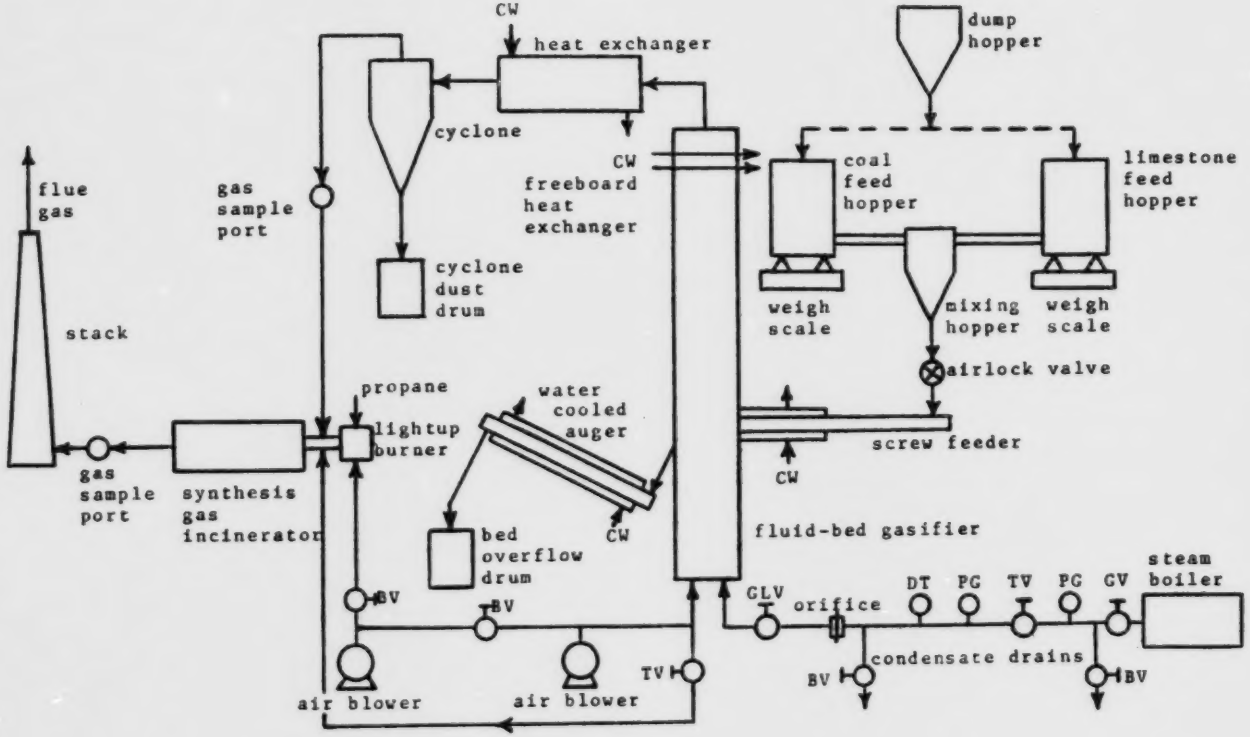


Fig. 3.1. Flow diagram of the modified pilot plant

Acronyms:

BV	ball valve	GV	gate valve
CW	cooling water	PG	pressure gauge
DT	dial thermometer	TV	throttle valve
GLV	globe valve		

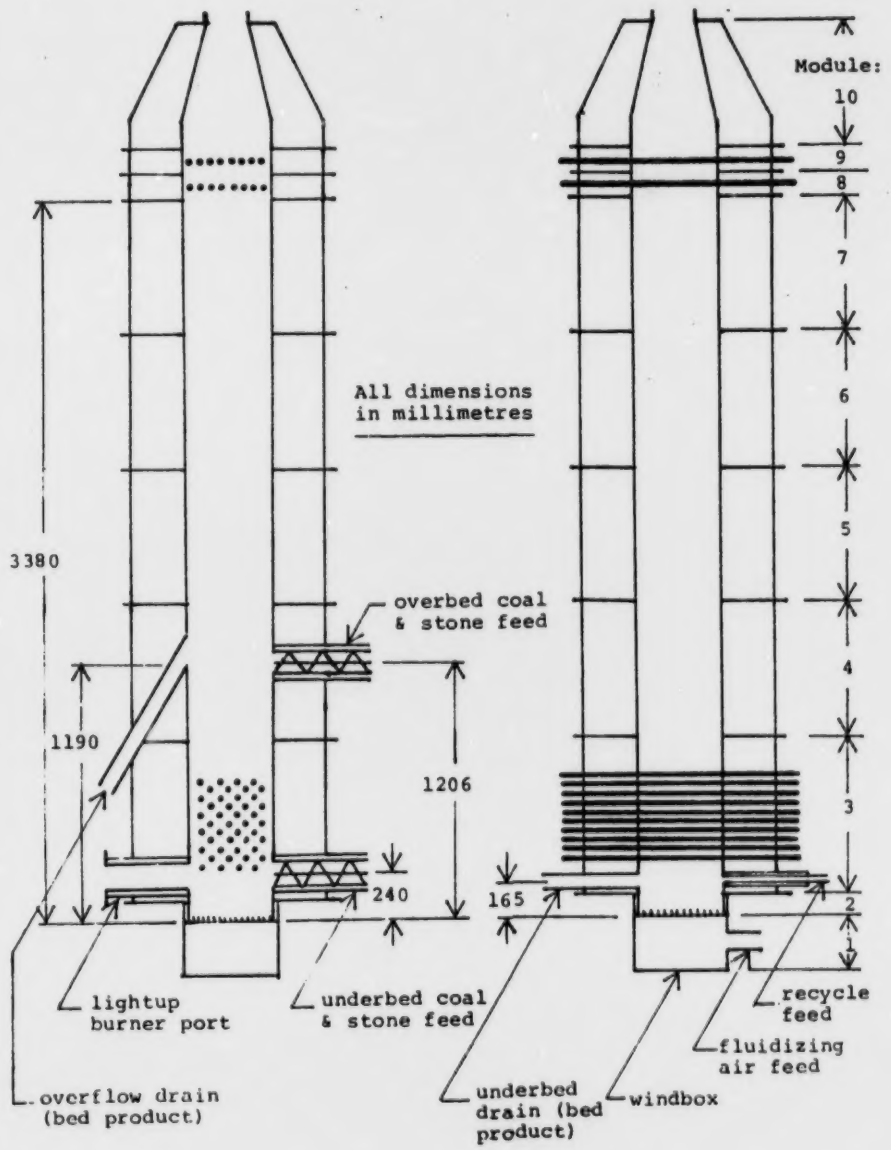


Fig. 3.2. Principal dimensions and features of the reactor

2. A waterwall downward extension, 102 mm high, which is clamped between modules 1 and 3. The addition of this extension increases the distance between the distributor plate and the underbed solids feedport, in order to improve the quality of fluidization and mixing at the solids feedpoint and to provide greater capacity for the accumulation of unfluidized solids before these interfere with the dispersion of the feed.

3. The in-bed heat exchanger module, 737 mm high. For the gasification studies, the heat exchange tubes (Fig. 3.2) were removed and the rectangular openings, normally occupied by the tubeport panels on opposite sides of the module, were filled with refractory brick and sealed by gaskets and steel plates bolted to the exterior surface of the module. This module also contains the ports (Fig. 3.2) for underbed feeding of coal and limestone, connection of the lightup burner, pneumatic injection of recycled cyclone dust, and underbed discharge of bed solids. The ports for cyclone dust recycle and underbed discharge were not used in the gasification trials.

4.--7. Four internally identical modules, each 635 mm high. The lowest has, Fig. 3.2, on opposite sides, two solids transport ports: a bed overflow port for the overflow of bed solids into a 77.9 mm ID (3-in, S-40) drain pipe set at 30° to the vertical, and an overbed feed port for coal and limestone. These modules, and module 3 as well, are all fitted with numerous ports for the insertion of sampling probes, thermocouples, and pressure taps.

Details are given in Becker and Code 1982.

8. & 9. Two crossflow freeboard heat exchanger modules, each 120 mm high and each containing a horizontal row of eight tubes carrying cooling water.

10. A transition module, 610 mm high, with a contraction at the end, leading into the exhaust ducting which carries the exhaust gases to the quench cooler, cyclone and synthesis gas incinerator. This module is fitted with ports for gas sampling. The external crossflow heat exchanger and the bag filter were bypassed in the gasification trials. A shell-and-tube heat exchanger surrounding the combustion chamber in the incinerator provided cooling for the exhaust gases prior to discharge to the stack. The bypass of the bag filter was a safety precaution taken to eliminate the potential hazard from the accumulation of a large volume of combustible gases in the bag filter.

Modules 3--10 are made up of 231 mm thick refractory brick contained in a steel shell 6.3 mm thick.

In summary:

- the gasifier has provision for overbed and underbed feeding.
- bed solids may be removed by overflow or by underbed extraction (not used in the gasification trials). These are the means of controlling the solids holdup, or the bed height.

- heat is extracted from the gases at the top of the freeboard by the freeboard heat exchanger.

- the height of the reaction chamber from the distributor plate at its bottom to the first row of tubes in the freeboard heat exchanger at the top, with the waterwall extension (module 2) in place, is 3.34 m.

- the bottom rim of the bed overflow port, with the waterwall extension in place, is 1.19 m above the distributor plate. This, then, is the nominal bed height when holdup is controlled by overflow.

3.4 The exhaust gas train

The exhaust gases from the gasifier pass through the following units on their way to the stack:

1. a short duct fitted with an explosion vent
2. the quench cooler
3. the cyclone dust separator, and
4. the synthesis gas incinerator.

The explosion vent, installed after the report by Becker and Code was issued, has a disk or lid firmly pressed over a 152 mm diameter hole by weights suspended on wires. The weights and wires are suitably sized so that in the event of an explosion the wires snap, the lid swivels up on a hinge, and the weights are transferred to restraining chains. The quench cooler and synthesis gas incinerator are described in the next subsection. A description of the redesigned windbox is also included in this subsection.

3.5 Modifications to the pilot plant

3.5.1 The quench cooler

A shell-and-tube heat exchanger was added to the system to rapidly quench the product stream leaving the top of the gasifier. The single pass tube bundle was assembled from 19, 5-ft (1.5-m) lengths of 1.5-in standard steel pipe (S-40; 48.3 OD; 40.9 mm ID) and enclosed in a 11.1-in (282-mm) ID cylindrical shell formed from 1/8-in (3.2-mm) steel plate. City water, in a single pass, is used as the coolant in the shell.

3.5.2 The synthesis gas incinerator

An incinerator was constructed to burn the synthesis gas produced in the gasifier. The unit consists of a central, cylindrical combustion chamber 35.7 cm in diameter and 1.5 m long surrounded by a two tube-pass, shell-and-tube exchanger. The lightup burner is also used as the pilot burner in the incinerator. When the fluid bed has reached a temperature sufficient for coal combustion, the lightup burner is removed from the gasifier and attached to a refractory lined tunnel extending from the incinerator to provide a hot jet of approximately 1950 °C along the axis of the combustion chamber. The synthesis gas is fed through an annulus surrounding the jet. Combustion air is supplied by diverting the excess air from the Roots blower, which is not required as feed to the gasifier, to a ring of dispersion jets in the annulus. The hot product gases

are cooled as they pass through the two concentric rings of tubes in a 2-pass arrangement surrounding the combustion chamber and exit to the stack. City water is used as the coolant in the shell.

In the modification of the pilot plant consideration was given to explosion hazards, particularly during the firing of the lightup burner when attached to the incinerator. In addition to the explosion vent in the discharge duct from the gasifier, explosion disks were installed in the duct connections at the outlets of the cyclone and the synthesis gas incinerator.

3.5.3 The windbox and distributor plate

A modified windbox and distributor plate for the injection of air and steam were designed and fabricated. The top and bottom surfaces of the waterwall extension (module 2 of the gasifier) contacting the interior flanges of module 3 of the gasifier and the top of the windbox were sealed with 'FIBERFRAX' rope. The connection was compressed by tightening the nuts on bolts attached to exterior flanges on module 3 and the windbox (module 1). The suspension of the windbox from the bottom of module 3 allowed easy removal of the windbox with a pallet truck for inspection of the interior of the gasifier and the removal of agglomerates and large unfluidizable stone particles from the bottom of the bed. The bottom chamber or steam chest was provided with a valved connection for the removal and measurement of condensate.

The distributor plate, which forms the top of the windbox, was fabricated from 0.5-in (12.7 mm) thick steel plate, 15.75 in by 14.75 in (400 mm by 375 mm). The 105 dispersion caps are arranged in 11 alternating rows of 10 and 9 caps in a triangular pitch. The centre-to-centre distances between caps are 1.548 in (39.32 mm) within rows and 1.341 in (34.06 mm) between rows. Five alternating caps in the 9-cap rows (25 total) are used for steam injection. A layout of the dispersion caps is illustrated in Fig. 3.3. An air bypass to the steam chest allows the steam caps to be used for air injection during startup of the unit. The caps were assembled from 1.5-in (38.1 mm) long sections of 0.5-in (S-80; 21.3 mm OD; 13.9 mm ID) and 7.35-in (186.7 mm) long sections of 0.25-in (S-80; 13.7 mm OD; 7.67 mm ID) heavy wall, 304 stainless steel pipe. The 0.25-in sections were inserted into the 0.5-in sections and the tops of the caps were sealed by welded 304 stainless steel rivets to form hemispherical heads. The bottom ends of the sections of 0.25-in pipe were threaded with 0.25-in NPT. Six, 1.4-mm diameter injection nozzles were drilled at 60° intervals through the walls of the concentric pipe sections, 7.6 mm below the base of the hemispherical heads. The location of the injection nozzles 30.5 mm above the plate provides an insulating layer of undisturbed solids for protection of the plate from the high temperatures in the bed. In addition, two 3/16-in (4.76 mm) diameter holes, located 2 in (5.1 mm) below the bottom of the 0.5-in pipe sections, were drilled on opposite sides of the 0.25-in pipe

sections of the 80 air injection caps to allow passage of air from the air chest to the injection nozzles. In the final assembly, the 0.5-in pipe sections of the caps rest on gasket washers on the surface of the plate and the longer 0.25-in pipe sections pass through holes drilled in the distributor plate and in the steel plate which forms the top of the lower chamber or steam chest. The threaded ends of the 0.25-in pipe sections which project into the steam chest were tightened and sealed by 'CAJON' brass pipe caps in the case of the air injection caps and fastened by hexagonal pipe nuts in the case of the steam injection caps. In the attachment of the dispersion caps, care was taken to position the nozzles so that the issuing jets are directed towards the spaces between adjacent caps.

SNC

3.6 The coal and limestone feed systems

The coal and limestone are charged through a dump hopper, located at floor level in the laboratory above the pilot plant, to the two feed hoppers mounted on the platforms of electronic weigh scales on the mezzanine level of the pilot plant. The metered feed streams from the coal and limestone hoppers enter a mixing hopper and are fed through a rotary air-lock valve ('star feeder') to an auger which discharges into the bed. The feed system may be operated in either overbed or underbed feed modes. More detailed descriptions of the equipment are given in the report by Becker and Code 1982.

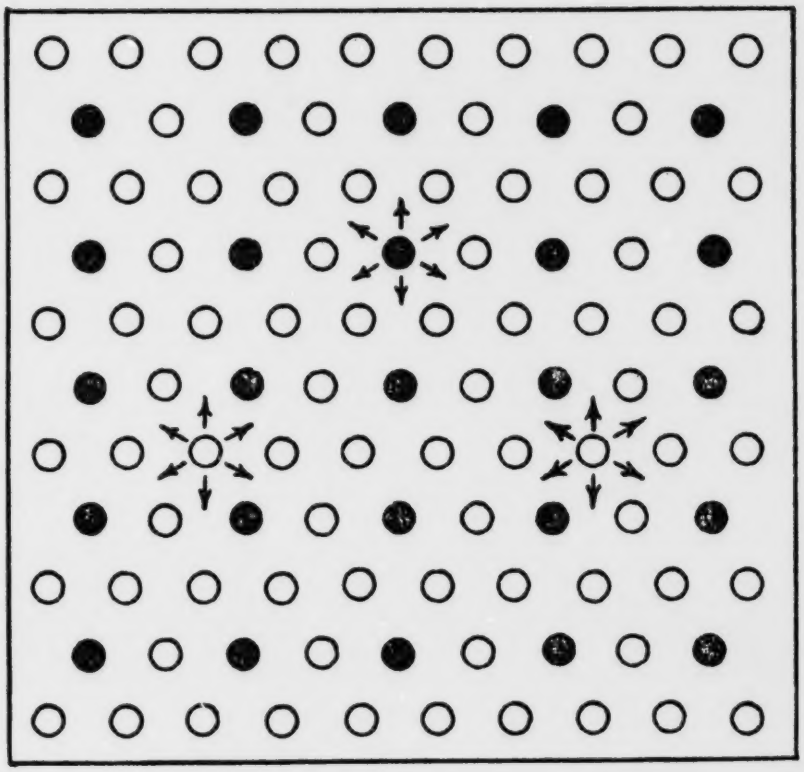


Fig. 3.3. Layout of dispersion caps on distributor plate

Key: ○ air dispersion cap; ● steam dispersion cap

3.7 Air supply and flows

Fluidizing air, fed to the windbox and dispersed through the distributor plate, is supplied by a Roots rotary lobe blower. Excess air from the blower, which is not required as feed to the gasifier, is diverted to the synthesis gas incinerator. Combustion air for the lightup burner is supplied by a second Roots blower. A small metered flow of air from the laboratory high-pressure air line is injected into the coal/limestone feed line to offset leakage through the rotary air lock valve and to block escape of bed gases through the valve. Similarly, a small metered purge of air is fed through the lines connecting the pressure taps of the steam orifice and the manometer in order to prevent condensation in the lines.

3.8 Steam supply system

Steam was required for the gasification work and a line to supply steam to the gasifier was installed between the steam chest of the windbox and the main from the department's auxiliary steam boiler. The insulated line of 2-in steel pipe (S-40; 52.5 mm ID; 60.3 mm OD) contains an orifice for flow measurement, a throttle valve for flow control, and a dial thermometer and pressure gauge to determine the condition of the steam entering the orifice. Condensate drains are located at the outlet of the steam main and just upstream of the orifice plate.

The orifice is a Meriam sharp-edged orifice with flange

taps. Interchangeable orifice plates with 0.750-in (19.1-mm) and 0.500-in (12.7-mm) diameter orifices are used depending upon the range of flowrates. The orifice pressure taps are connected to the legs of a U-tube manometer by teflon tubing. Water was selected as the manometer fluid for compatibility with steam condensate. Surge tanks, installed in the tap lines to damp out pressure oscillations, also act as condensate knockout drums. A small metered purge of air is injected into the tap lines to sweep steam from the lines.

The flowrate of steam, indicated by the manometer reading, is controlled manually by adjusting the 0.5-in throttle valve. The controls on the boiler were set to produce saturated steam at 90 psig (620 kPa gauge). This steam pressure is sufficient to offset heat losses and to ensure, following throttling through the insulated throttle valve, that the steam is superheated, i.e., that the steam passes through the orifice as a single phase fluid. The dial thermometer and pressure gauge, located just upstream of the orifice, are monitored to determine the condition of the steam and hence the specific volume which is required in the calculation of the steam flowrate.

3.9 Instrumentation and analytical systems

The gas sampling and analysis systems, the temperature and pressure measurement systems, and the data acquisition and processing system (based on a desk-top minicomputer,

Hewlett-Packard Model 9836) are described in the report by Becker and Code 1982. In summary, infrared analyzers are used for on-line measurement of CO, CO_2 and SO_2 concentrations, a paramagnetic analyzer for O_2, a chemiluminescence analyzer for NO and NO_x, and a flameionization analyzer for unburnt hydrocarbon ('HC', indicated as ppm of CH_4). Thermocouples, with backup dial thermometers at critical points in the air and water circuits, are used for temperature measurement. The above instrumentation for combustion gas analysis was in place. The gas chromatograph (Hewlett-Packard Model 588Ø) from the analytical laboratory was installed in the pilot laboratory for analysis of the synthesis gas. Gas sample ports and probes were installed in the exit duct from the cyclone and in the exhaust duct from the incinerator for the withdrawal of samples of synthesis gas and combustion gas, respectively. The samples of synthesis gas were analyzed wet. When a dry analysis was required the sample was dried in the drying aparatus in the combustion gas analysis train (Becker and Code 1982).

Proximate analysis and sulfur analysis of the coal feed and samples of the cyclone product and bed material were carried out in our analytical laboratory. Selected samples of the solids were analyzed for carbon, hydrogen, nitrogen and specific sulfur-containing species by a commercial laboratory.

4. MATERIALS: COAL AND LIMESTONE

4.1 Coal

4.1.1 General characteristics

An evaluation of Nova Scotia and New Brunswick coals has been made by Tibbetts and Montgomery (1976). Minto coal (central New Brunswick) is ranked as high volatile A bituminous (ASTM classification). It is a fairly strongly swelling coal with a free swelling index (FSI) range of 5 to 5.5. Ash content is in the range of 15--25 %. The sulfur content of Minto coal is typically 7--8 %.

4.1.2 Origin and handling of Minto Coal

The Minto coal was purchased from a mine operator (New Brunswick Coal Ltd., Minto, N.B.) as run-of-mine coal, containing lumps up to 20 cm diameter, hauled to Kingston, and crushed (in a roll mill) and screened by a Kingston area quarry operator. The coal used in the gasification work was a minus 3/16-in fraction.

4.1.3 Chemical analyses

Proximate analyses and sulfur analyses were performed in our analytical laboratory. Carbon, hydrogen, nitrogen (CHN) analyses and analyses for specific sulfur-containing species were carried out by a commercial laboratory (referred to as analyst A in the material to follow). The moisture content (dry basis) of the Minto coal was 1.9 ± 0.8 %. This is an average value based

on the results from 9 samples. The result obtained by analyst A on a sample of the coal feed used in run 830705 was 2.0 %. Other results of the proximate analyses are summarized in Table 4.1. The ratio of fixed combustible matter (often referred to as 'fixed carbon') to total combustible matter, FCM/CM, based on the analyses of the 9 samples referred to above, was 0.543. This value is lower than the value of 0.600 reported by Tibbetts and Montgomery (1976) for a typical run-of-mine sample of Minto coal. A value in closer agreement, 0.594, was obained in the earlier combustion studies on Minto coal (Becker and Code, 1983).

Ultimate analyses are summarized in Table 4.2, again with comparisons with published results. The total sulfur content of 6.4 % reported by analyst A for a sample of the coal feed from run 830705, although lower than the composite value and the value given by Tibbetts and Montgomery (1976), is in agreement with our analytical result, 6.3 ± 0.1 %. This average value is based on the analyses of nine samples. It is our experience that the sulfur content is lower in small sized fractions than in run-of-mine samples. The oxygen contents, obtained by difference, should be regarded with reservation. The ash, for example, presumably contains any sulfate sulfur originally present in the coal. The sulfur analysis, however, gives the total sulfur. Thus the sulfate sulfur is counted twice. The analysis for sulfur forms in Minto coal reported by analyst A gives a value of 0.76 % (dry basis) for sulfate sulfur (Table 4.3). Thus this sample of Minto coal contains 0.76 % sulfate

Table 4.1. Proximate analyses of Minto coal, dry basis

Reporter	Queen's[a]	σ	Queen's[b]	T & M[c]
VCM, %	38.6	1.7	33.7	32.5
FCM, %	45.9	2.3	49.3	48.8
Ash, %	15.5	1.3	17.4	18.8
FCM/CM	0.543		0.594	0.600

[a] This work, based on the analyses of nine samples, - 3/16-in fraction, $\hat{D}_p$, 2.2 mm

[b] Becker and Code (1983), average of nine analyses, three samples each from three particle size ranges: - 1 1/2 in + 7/8 in, $\hat{D}_p$ 19.3 mm; - 7/8 in + 3/8 in, $\hat{D}_p$ 10.7 mm; - 3/8 in + 3/16 in, $\hat{D}_p$ 6.6 mm

[c] Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL No. 2073-76)

Note:
VCM, volatile combustible matter FCM, fixed combustible matter
CM = VCM + FCM, total combustible matter

$\hat{D}_p$, characteristic diameter for particles of mixed size, defined as the diameter corresponding to a mass fraction undersize of 0.5, based on sieve analysis.

Table 4.2. Ultimate analysis of the coal dry matter. Oxygen, O, is estimated by difference.

Reporter	Analyst A[a]	Composite[b]	T & M[c]
C, %	67.7	68.3	66.3
H, %	4.7	4.1	4.3
N, %	0.7	0.8	0.9
S, %	6.4	7.4	8.0
Ash, %	17.9	17.4	18.7
O, %	2.6	2.0	1.8

[a] Sample of coal feed used in run 830705

[b] Becker and Code (1983), composite values based on analyses of samples of Minto coal by several outside analysts

[c] Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL No. 2073-76)

Table 4.3. Sulfur forms in Minto coal

Pyritic sulfur,	%	3.32
Sulfate sulfur,	%	0.74
Organic sulfur,	%	2.17
$S^{=}$,	%	0.05
Total sulfur ,	%	6.28

Analysis, on an as received basis, by analyst A on a sample of coal feed from run 830705

sulfur and 5.65 % non-sulfate sulfur. In the case of Minto coal this has a significant effect on the estimate of the oxygen content. An additional consequence of this procedure is that all of the errors in the ultimate analysis are combined in the estimate of the oxygen content.

4.1.4 Ash fusibility

The ash of Minto coal, Table 4.4, softens at relatively low temperatures. The softening of the ash has an adverse effect on the quality of fluidization and mixing in the bed. This was our experience in the gasification of Minto coal. The agglomeration of the coal particles resulted in poor dispersion, the development of local zones of high temperature and subsequent loss of temperature control in the bed. The temperatures characterizing the fusibility of Minto coal in Table 4.4 are for oxidizing conditions. Corresponding temperatures for Devco coal are included in Table 4.4 for both oxidizing and reducing conditions. The temperatures under reducing conditions are lower by 170 °C from the values under oxidizing conditions. This would lead one to suspect that operating problems with Minto coal under gasification (reducing) conditions would develop at significantly lower temperatures than the values given in Table 4.4.

4.1.5 Calorific value

The calorific value of Minto coal was determined in our laboratory using a Parr combustion calorimeter. Based on 25 determinations, the value on an ash-free, moisture-free basis

Table 4.4. Ash fusion characteristics for Minto and Devco coals

Coal	Minto[a]	Devco[b]	Devco[b]
Reporter	T & M	CBDC	CBDC
Conditions	Oxidizing	Oxidizing	Reducing
IT, °C	1066	1259	1091
ST, °C	1099	1319	1141
HT, °C	1138	1368	1186
FT, °C	1215	1406	1310

[a] Tibbetts and Montgomery (1976), typical run-of-mine sample (ERL No. 2073-76)

[b] Data for Devco coal (Prince Mine) furnished by Cape Breton Development Corporation, based on product over one week in 1981

Note:

IT, initial temperature
ST, spherical softening temperature
HT, hemispherical softening temperature
FT, fluid temperature

(calorific value of the combustible matter) was 34.6 ± Ø.4 MJ/kg. This is in close agreement with the value of 34.9 ± Ø.3 MJ/kg given by Tibbetts and Montgomery (1976) for ten samples from the Minto field.

4.1.6 Particle size analyses

Particle size analyses were carried out by sieve analysis on two samples of the coal feed, run 830627 and run 830712. The results, Fig. 4.1, for the two samples are in good agreement except near the tails of the distributions which approach a log-normal distribution over the particle size range Ø.2 mm < D_p < 4 mm, corresponding to 3 % and 8Ø % undersize. A characteristic particle diameter, $\hat{D}_p$, defined as the particle diameter corresponding to 5Ø % by mass undersize, is 2.2 mm.

4.2 Limestone

4.2.1 Chemical Analysis

Table 4.5 shows the chemical analysis of the present lot of Havelock limestone (Becker and Code, 1983). Also shown, for comparison, is an analysis of limestone from the same quarry performed at the Mineral Sciences Laboratory, CANMET, Energy Mines and Resources Canada. Assuming that the loss on ignition is all CO_2, the results for the present lot of stone indicate a CO_2 content of 9.67 mol/kg. The element analysis indicates 9.68 mol/kg of Ca and Ø.Ø9 mol/kg of Mg, or a combined Ca and Mg content of 9.77 mol/kg, in fairly good agreement with the molar CO_2 content. Our estimate of the effective $CaCO_3$ content of the

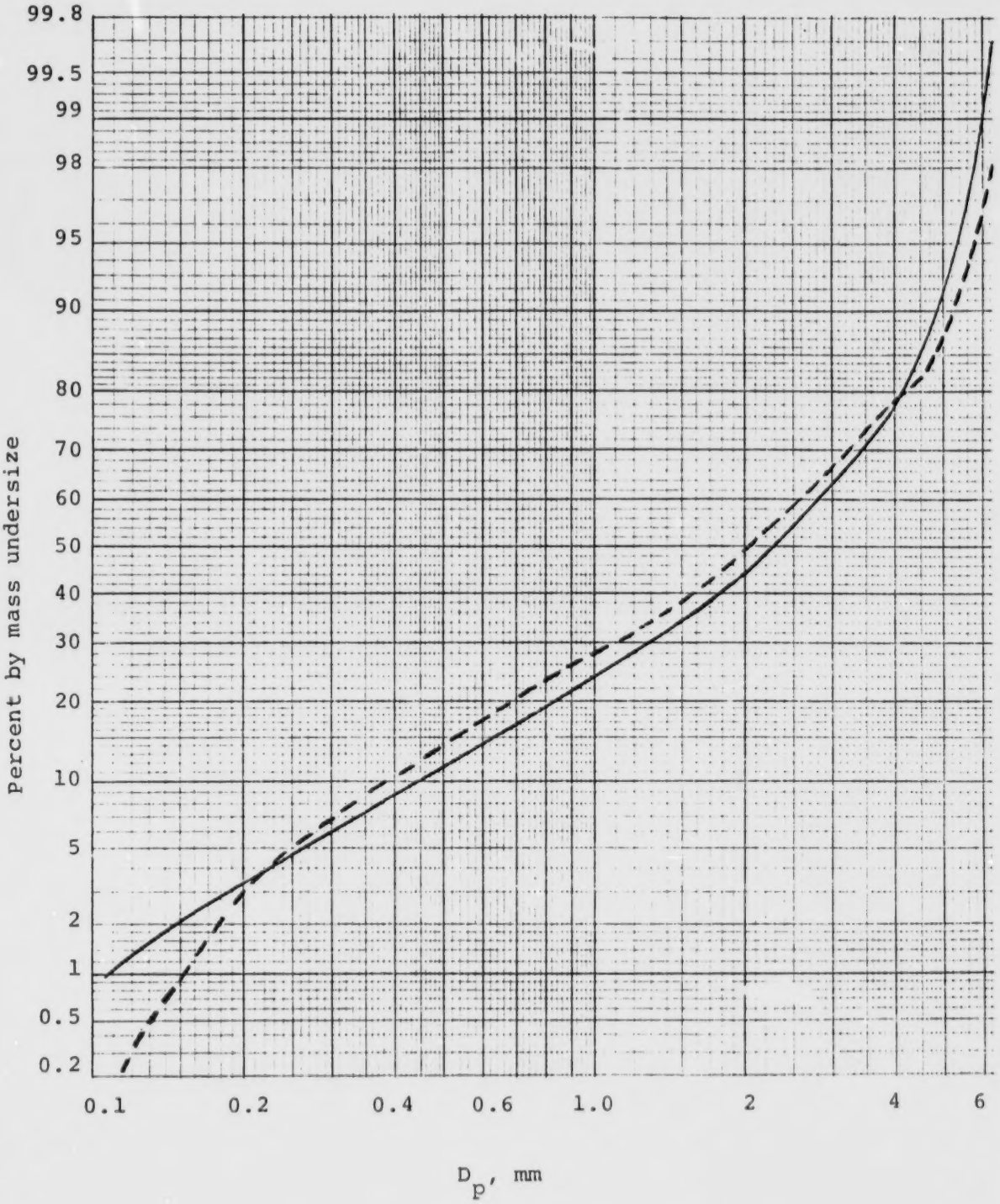


Fig. 4.1. Particle size distribution of coal feed.

--- Run 830627 —— Run 830712

Table 4.5. Analysis of dry Havelock limestone

Composition by elements			Composition by oxides		
Element	% (QFBC)[a]	% (MSL)[b]	Oxide	% (QFBC)[a]	% (MSL)[b]
Al	0.50	0.08	Al_2O_3	0.94	0.15
Ca	38.8	39.52	CaO	54.3	55.33
Fe	0.37	0.12	Fe_2O_3	0.53	0.17
Mg	0.22	0.41	MgO	0.36	0.68
Si	0.44	0.63	SiO_2	0.94	1.35
S	--	< 0.01	SO_3	--	< 0.03
Total oxides				57.07	57.68
Loss on ignition at 980 °C				42.57 ± 0.65	42.60
Oxides plus ignition loss				99.64	100.28

Other MSL results:

Est. Ca present as $CaO + Ca(OH)_2$: 0.05 %

Est. Mg present as $MgO + Mg(OH)_2$: < 0.01 %

Est. Ca present as $CaSO_4$: < 0.01 %

[a] Becker and Code (1983). Element analyses by outside analyst B. Loss on ignition obtained in our laboratory (mean in scores of runs in our bench-scale studies of calcination kinetics)

[b] Analysis by Mineral Sciences Laboratory, CANMET, Energy Mines and Resources Canada (their sample No. MS-CL-78-327)

limestone is 96.8 %. This estimate was obtained by taking the above value of 9.67 mol/kg which is based on a large number of determinations in our bench-scale calcination kinetics and lumping Ca and Mg, weighted according to the element analysis.

4.2.2 Particle size analysis

The particle size distribution obtained by sieve analysis of a sample of the limestone feed (run 830712) is shown in Fig. 4.2. The distribution is approximately log-normal for the particle size range 0.1 mm < D_p < 0.8 mm, corresponding to 3 % to 80 % undersize. The characteristic particle size $\hat{D}_p$, as defined for the coal, is 0.48 mm.

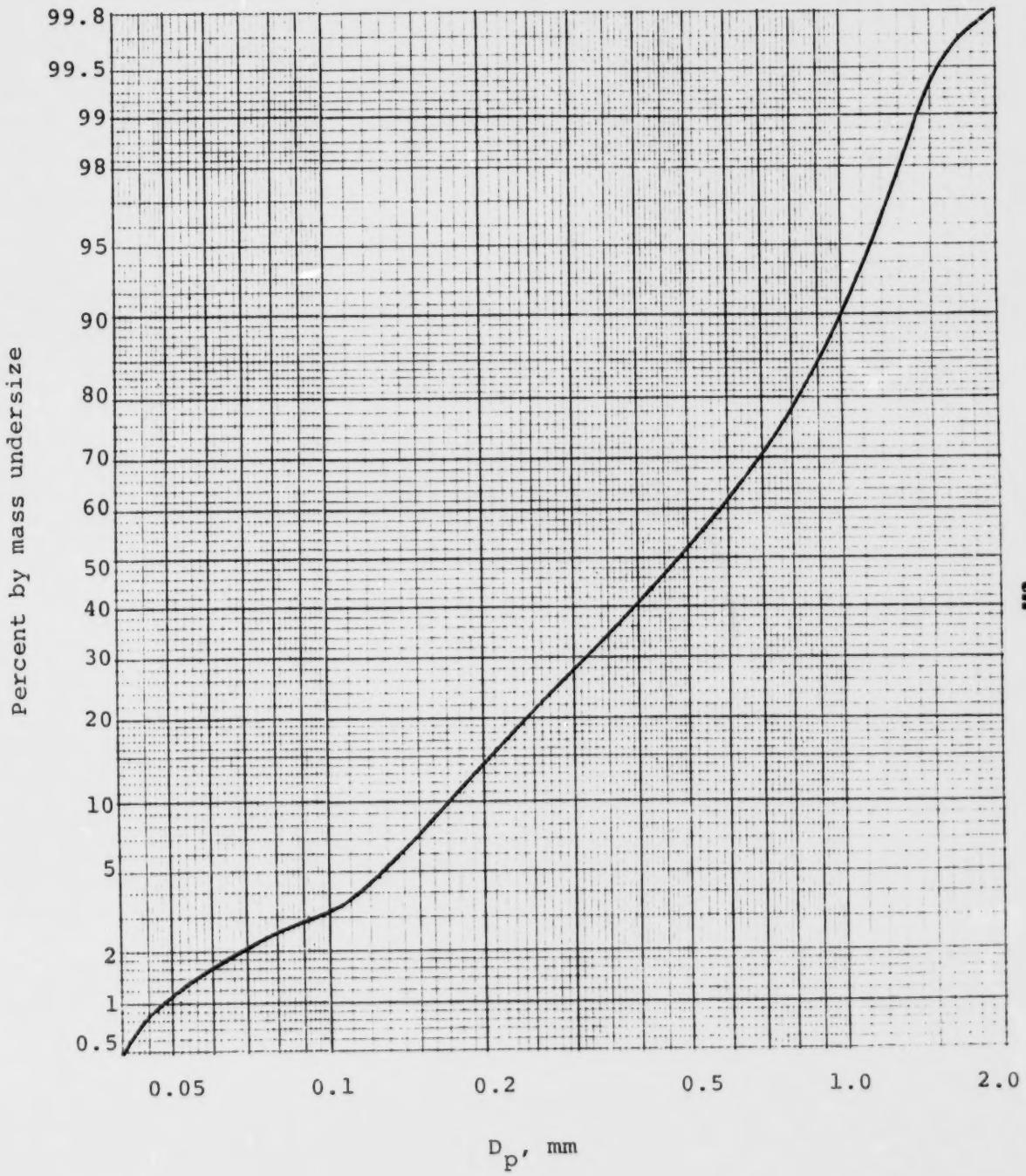


Fig. 4.2. Particle size distribution of limestone feed

5. EXPERIMENTAL TESTS

5.1 Operating procedures

5.1.1 Introduction

The gasifier was operated as a combustor during the startup period. When the bed temperature approached the operating temperature, the system was switched over to gasification conditions by establishing the steam flow and rapidly increasing the flow of coal and limestone. The endothermic energy required for gasification was supplied by partial oxidation of the coal. The control strategy adopted was to maintain the fluidization velocity and the steam-to-air ratio at the set values and to control the bed temperature by adjusting the coal feed rate. The limestone feed rate was concurrently adjusted to maintain the Ca/S ratio at the set value. The agglomeration of particles in the bed required the dropping of the windbox at the end of a run to remove the bed material and clean the agglomerates from the walls of the gasifier and the distributor plate. Thus each run started from cold conditions.

5.1.2 Cold startup

The procedure for startup from cold conditions was as follows:

1. Move the mobile lightup burner into the pilot laboratory, connect service (air, propane, electrical) lines and attach to

the burner port on the gasifier.

2. Charge the gasifier with bed material from a previous run to around 80 % of the expected steady-state holdup. Screen the bed material prior to charging to remove large particles. If no appropriate bed material is available charge the bed with limestone.

3. Establish cooling water flows to all heat exchangers and cooling jackets.

4. Establish a fluidizing airflow sufficient to give good agitation, typically a fluidizing velocity under cold conditions of around 0.4 m/s (corresponding to 1.5 m/s at 850 °C).

5. Establish a flow of purge air to the solids feedline (air-lock valve).

6. Start the lightup burner at low fire. Increase the firing rate at bed temperatures of 150 °C and 300 °C.

7. Commence feeding coal and limestone at a bed temperature of 600 °C.

8. When the bed reaches a temperature of 850 °C, slump the bed (turn off the fluidizing air and solids feed), quickly disconnect the lightup burner, cover the burner port with a blind flange and

move and attach the lightup burner to the synthesis gas incinerator.

9. Attach the bed-product cooler (water-cooled auger) to the bed solids overflow port and attach a product collection barrel to the auger discharge.

10. Restart the lightup burner in its position as the pilot burner of the synthesis gas incinerator.

11. Restart the flow of fluidizing air at the set operating level and recommence feeding coal and limestone.

12. When the bed reaches a temperature in the range 800 to 850 °C, establish the flow of steam to the steam chest of the windbox at the set operating level. The level of steam flow is determined from the set fluidizing velocity (air plus steam) and the specified steam-to-air ratio for the run. Before opening the throttle valve, purge accumulated condensate from the steam line by cracking the valve on the drain line upstream of the throttle valve. Repeat this step periodically throughout the run at the two drain lines, upstream of the throttle valve and upstream of the orifice plate. Close the valve on the air bypass to the steam chest.

13. Establish gasification conditions by rapidly increasing the feed rates of coal and limestone. Control the bed temperature by

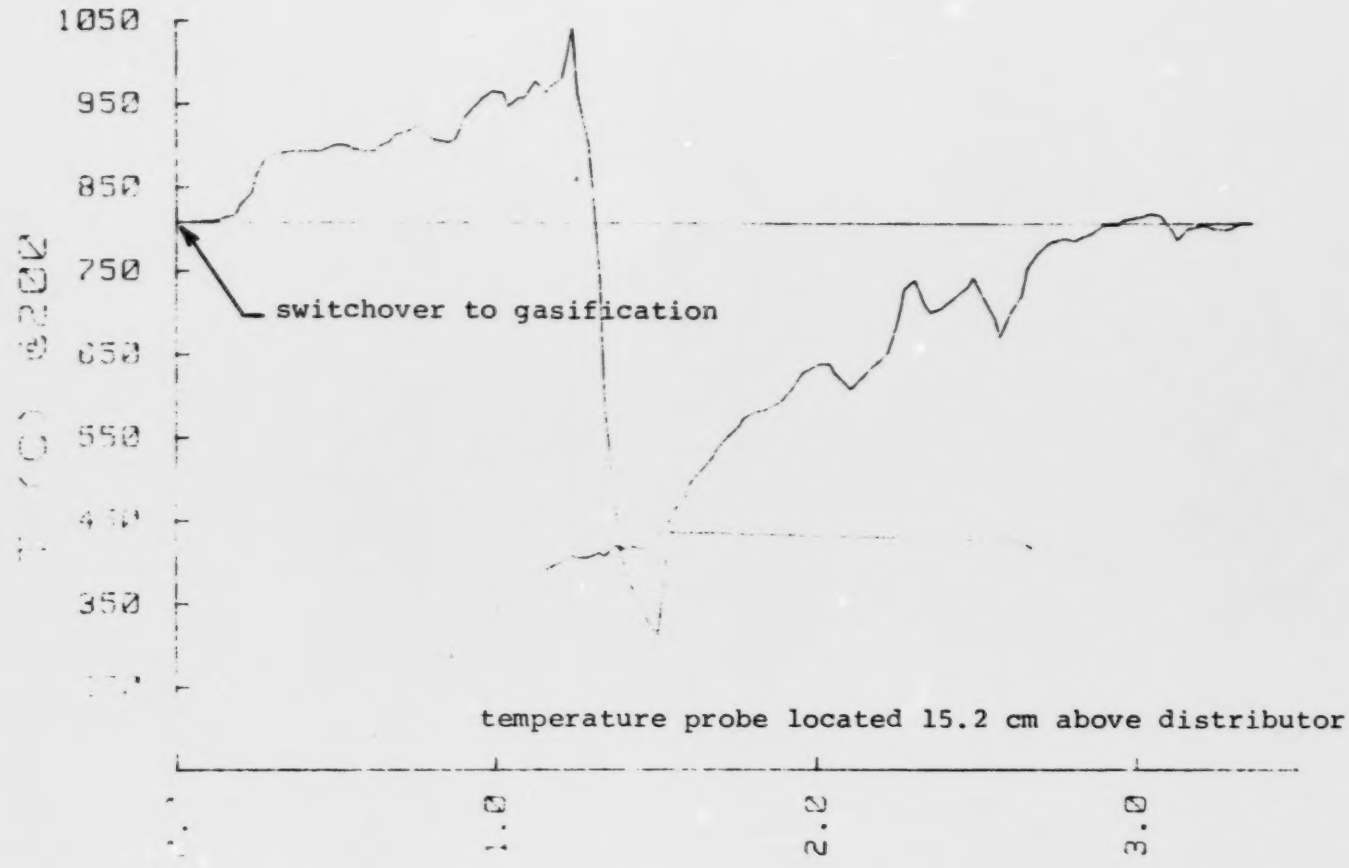
adjusting the feed rates of coal and limestone while maintaining the Ca/S ratio at the specified value. Note that it was not always possible to maintain the Ca/S ratio at the specified value during this unsteady adjustment period. The limestone feed was used frequently as a quench during the establishment of gasification conditions.

14. When the bed depth reaches its operating level, as indicated by the bed pressure drop, open the valve on the bed-solids discharge port and commence removing bed material to maintain constant bed depth. In the gasification work, bed depth was regulated by overflow, using the overflow port located 1.2 m above the distributor plate.

15. During startup and in subsequent operation, adjust all cooling water flows so exit temperatures are no higher than 60 °C. This rule ensured the avoidance of boiling.

The time required to attain gasification conditions was typically 3 to 3.5 hr. Fig. 5.1 shows a computer trace of the temperature at a bed level 15.2 cm above the distributor plate. The time was reset to zero (Fig. 5.1) at the point of switchover to gasification conditions. Higher rates of temperature rise are attainable but detrimental to the refractory walls of the gasifier.

Fig. 5.1. RUN 830707 BED TEMPERATURE (C)



5.1.3 Procedure following startup

Full data collection generally began with the start of bed-solids removal. The following variables were recorded at suitably close intervals:

- o weights of the coal and limestone feed hoppers, to determine the feed rates of coal and limestone.

- o air temperature and pressure drop (manometer reading) across the air orifice, to determine the air flowrate.

- o upstream temperature and pressure of the steam and pressure drop (manometer reading) across the steam orifice, to determine the steam flowrate. The accumulated condensate in the steam chest was measured at hourly intervals and the steam flowrate was adjusted to compensate for condensation in the steam chest. This correction amounted to only a few percent of the steam metered.

- o effluent production of bed solids and cyclone dust. The collection barrels on the discharges of the bed-solids cooler and the cyclone were replaced with empty barrels and the masses of bed material and cyclone dust collected were determined by weighing. Samples of the solids were collected and identified for later analysis. There was no recycle of cyclone dust.

- o gas composition at the exit from the cyclone. Depending upon whether a wet or dry analysis was desired, the sample of synthesis gas was analyzed directly or dried in the drying apparatus in the combustion gas analysis train prior to injection into the gas chromatograph.

- o temperatures in the bed and freeboard and at other critical points in the system as indicated by fixed thermocouple probes.

- o static pressures in the windbox, at several levels in the bed, and at other significant points in the system.

The instrumentation for combustion gas analysis, interfaced with the computer, was used occasionally for the continuous analysis of the stack gas over three minute intervals.

5.1.4 Shutdown procedure

The shutdown procedure followed a sequence of the steps somewhat in reverse of those used in startup.

1. Shut off the coal feed.

2. Shut off the limestone feed. The interval of time between the shutoff of the coal and the shutoff of the limestone depended upon the temperature behavior of the bed. The limestone is a

convenient quench (sensible heat effect plus endothermic calcination reaction) to rapidly decrease the bed temperature.

3. Turn off the steam flow and shut down the steam boiler. Increase the air flow to the bed and open the valve on the air bypass to the steam chest to allow the dispersion of air through the steam nozzles.

4. After a sufficient interval of time to ensure that all combustibles have been expelled from the system shut off the pilot burner on the synthesis gas incinerator.

5. Shut off the solids feed auger and the star feeder (air-lock valve).

6. Close the valve on the bed overflow port, detach the bed-solids cooler, disconnect the water line to the cooling jacket on the auger and move the bed-solids cooler out of the pilot laboratory to allow access to the west face of the gasifier.

7. When the bed temperature has dropped below 200 °C turn off the air blower and the purge of laboratory air to the solids feedline.

8. Remove any residual steam and condensate in the steam line, between the throttle valve and the orifice, by repetitive pressurization of the line with laboratory air (fed through the

crifice taplines) and discharge through the drain line.

9. Set the cooling water flows at moderate rates and leave on over the period that the bed is slumped in order to avoid boiling if localized heating develops.

5.1.5 Post-shutdown work

Much of the downtime between runs was taken up refurbishing the gasifier. Before removing the solids from the bed, the air blower was turned on and the fluidization air was set at a moderate flowrate to allow a check to be made of the bed temperature and to provide cooling for any hot spots that may have developed during the slump period. With the bed under a low level of agitation, the bulk of the solids were removed with a hand-operated auger inserted into the bed through the solids withdrawal port. The air and steam lines to the windbox and the water line to the waterwall extension were disconnected and with the aid of a pallet truck the windbox was dropped and removed to allow access to the interior of the gasifier for removal of agglomerates from the walls. Agglomerates were removed from around the dispersion caps and the distributor plate was cleaned with a shop vacuum cleaner. The windbox was reattached and sealed to the base of the gasifier and the air, steam and water lines were reconnected. The bed solids were screened to provide bed material for the next run and the feed hoppers were charged with coal and limestone.

5.2 Test program

The operating conditions specified for the planned series of preliminary runs are summarized in Table 5.1. Operating difficulties presented by excessive carryover of solids and particle agglomeration in the bed required an extensive revision of our original plan. A summary of the operating details for the runs undertaken is given in Table 2.1.

A fluidizing velocity of 2.3 m/s, used in the first five runs, resulted in excessive carryover of solids and plugging of the tubes in the overhead exchanger and duct to the cyclone. This problem was overcome by reducing the fluidizing velocity (1.3 m/s in the next five runs and 1.6 m/s in the final six runs).

Agglomerate formation in the bed presented a much more difficult problem to overcome and the test program concentrated on the solution of this problem. The use of a dry coal feed and different limestone-to-coal ratios had no significant effect on the rate of agglomeration. The effect of agglomerate formation on the temperature distribution in the bed is illustrated in Fig. 5.2 through Fig. 5.7 where the thermocouple readings at two levels in the bed, 15.2 cm and 54.7 cm above the distributor plate, are plotted against the time of operation at gasification conditions. Temperature problems due to particle agglomeration are indicated by an increasing disparity between the thermocouple readings at the two levels followed by an apparent rapid decrease

Table 5.1. Operating conditions specified for the preliminary set of tests

$W_{a,c}$	$W_{s,c}$	$W_{s,a}$	X_s	x_s	$\dot{n}_a$	$\dot{n}_s$	$\dot{m}_a$	$\dot{m}_c$	Ca/S	1.5	3.0
air/coal	steam/coal	steam/air	steam/air	steam/gas in					$\dot{m}_g/\dot{m}_c$	0.39	0.70
kg/kg	kg/kg	kg/kg	mol/mol	mol/mol	kmol/hr	kmol/hr	kg/hr	kg/hr		$\dot{m}_g$, kg/hr	$\dot{m}_g$, kg/hr
2.1	0.2	0.095	0.153	0.133	11.7	1.80	339	161		63	113
2.1	0.5	0.238	0.383	0.277	9.8	3.74	284	135		53	95
3.7	0.2	0.054	0.087	0.080	12.4	1.08	360	97		38	68
3.7	0.5	0.135	0.218	0.179	11.1	2.42	322	87		34	61

$\dot{n}_g = \dot{n}_a + \dot{n}_s = 13.5$ kmol/hr

$T_b = 900$ °C

$v_f = 2.3$ m/s

Fig. 5.2

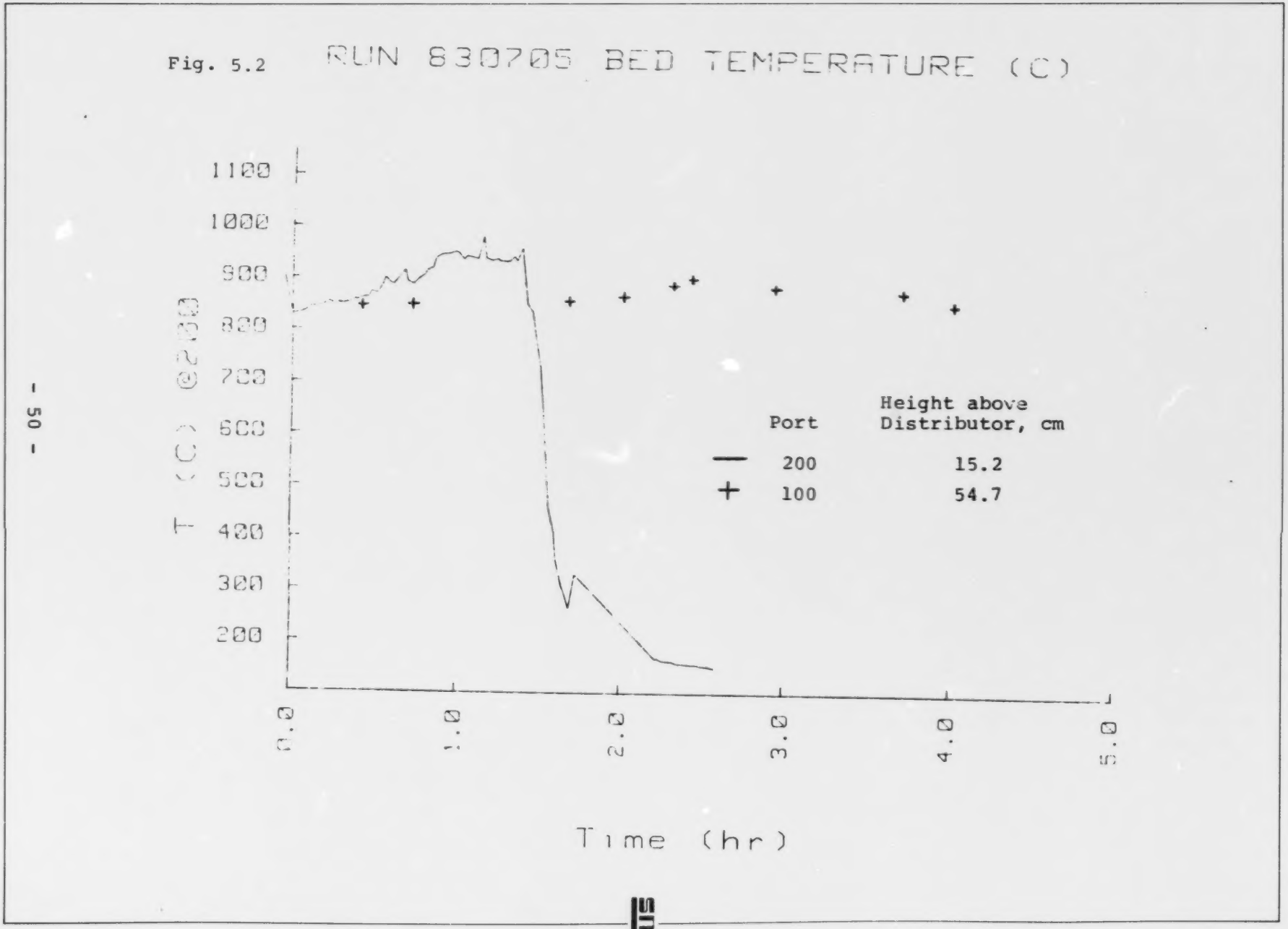
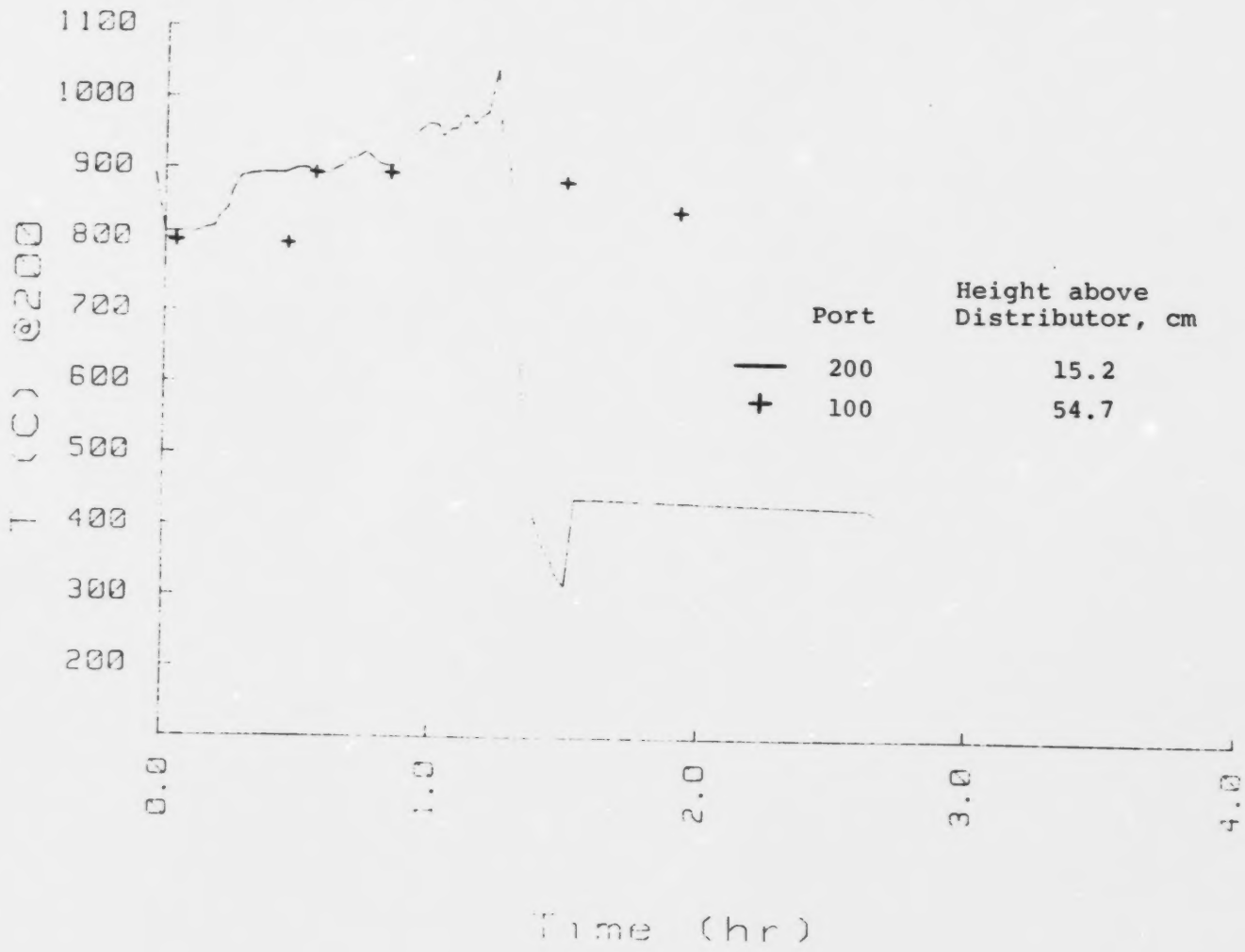

RUN 830705 BED TEMPERATURE (C)
T (C) @2.00
1100
1000
900
800
700
600
500
400
300
200
0.0
1.0
2.0
3.0
4.0
5.0
Time (hr)
Port
Height above Distributor, cm
200
15.2
100
54.7

Fig. 5.3 RUN 890707 BED TEMPERATURE (C)



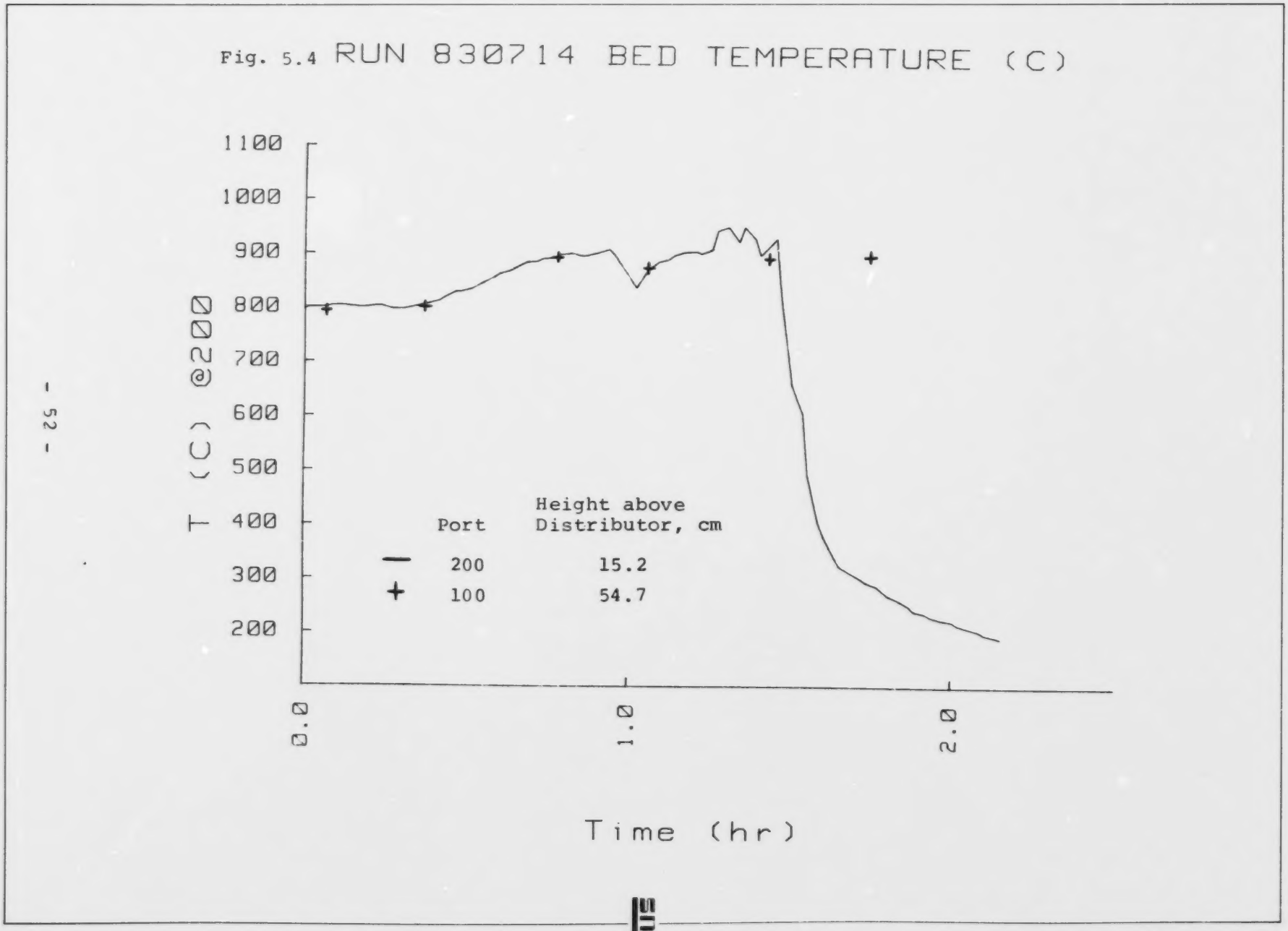


Fig. 5.4 RUN 830714 BED TEMPERATURE (C)
1100
1000
900
800
700
600
500
400
300
200
T (C) @200
Port
Height above Distributor, cm
200
15.2
100
54.7
0.0
1.0
2.0
Time (hr)

Fig. 5.5 RUN 830712 BED TEMPERATURE (C)

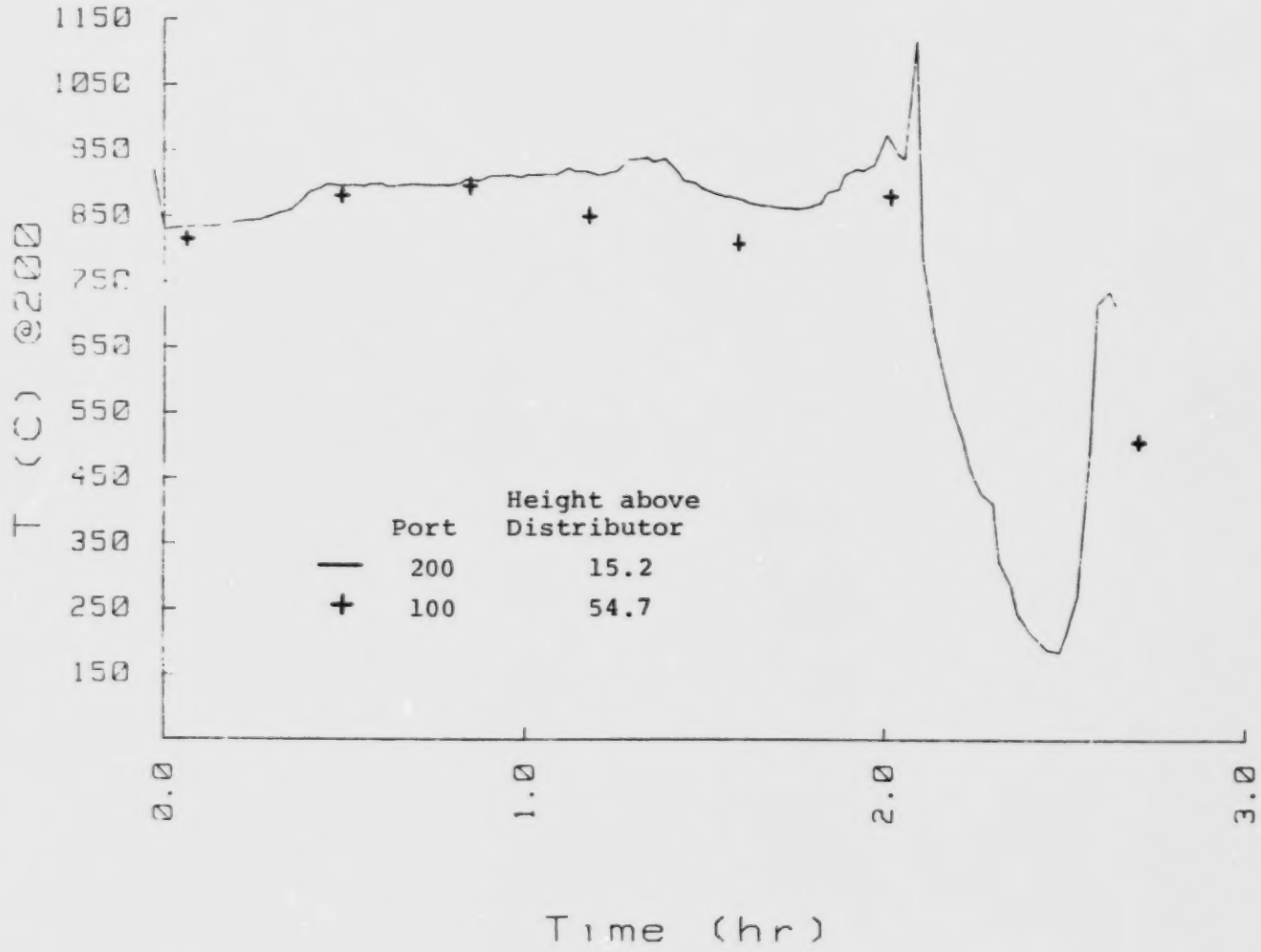


Fig. 5.6 RUN 83062? BED TEMPERATURE (C)

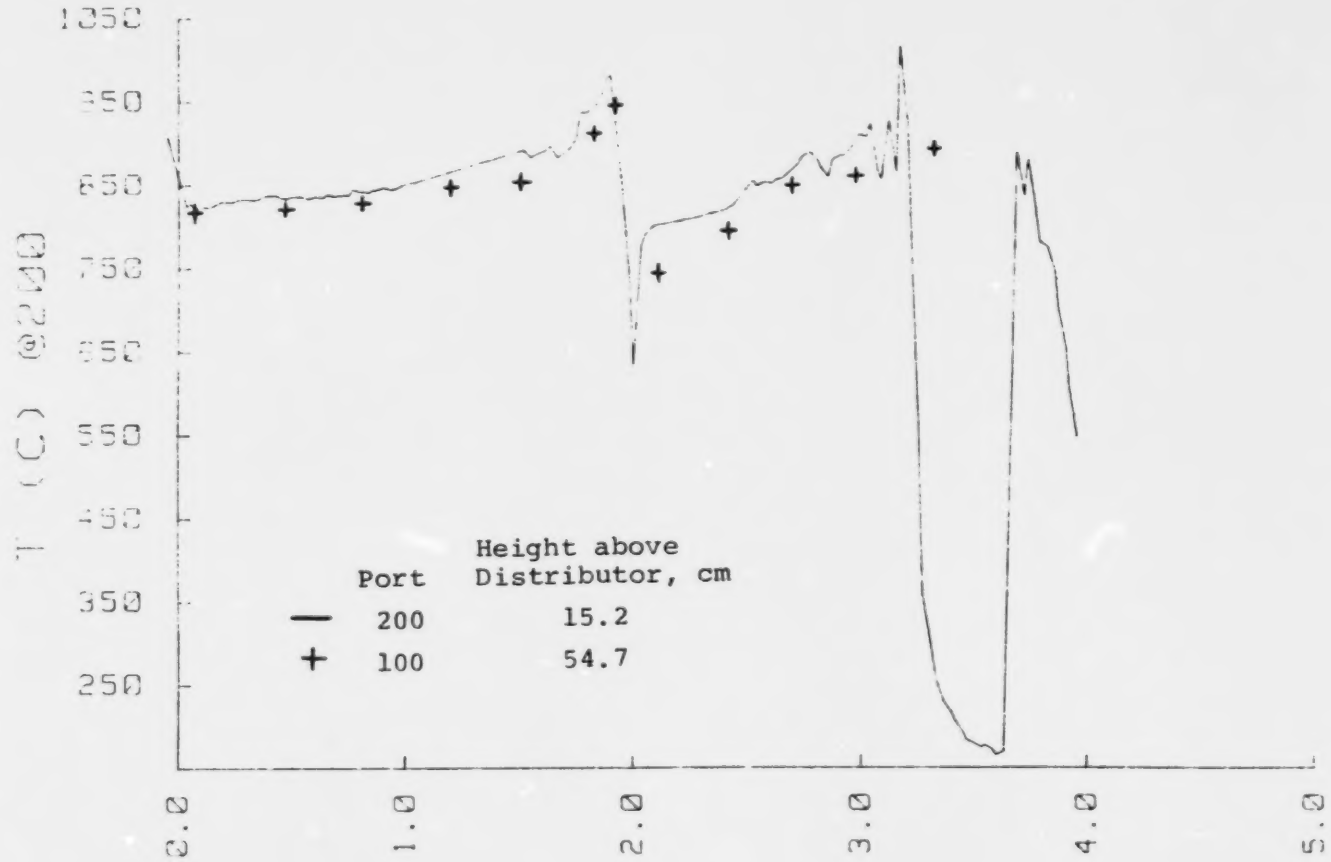
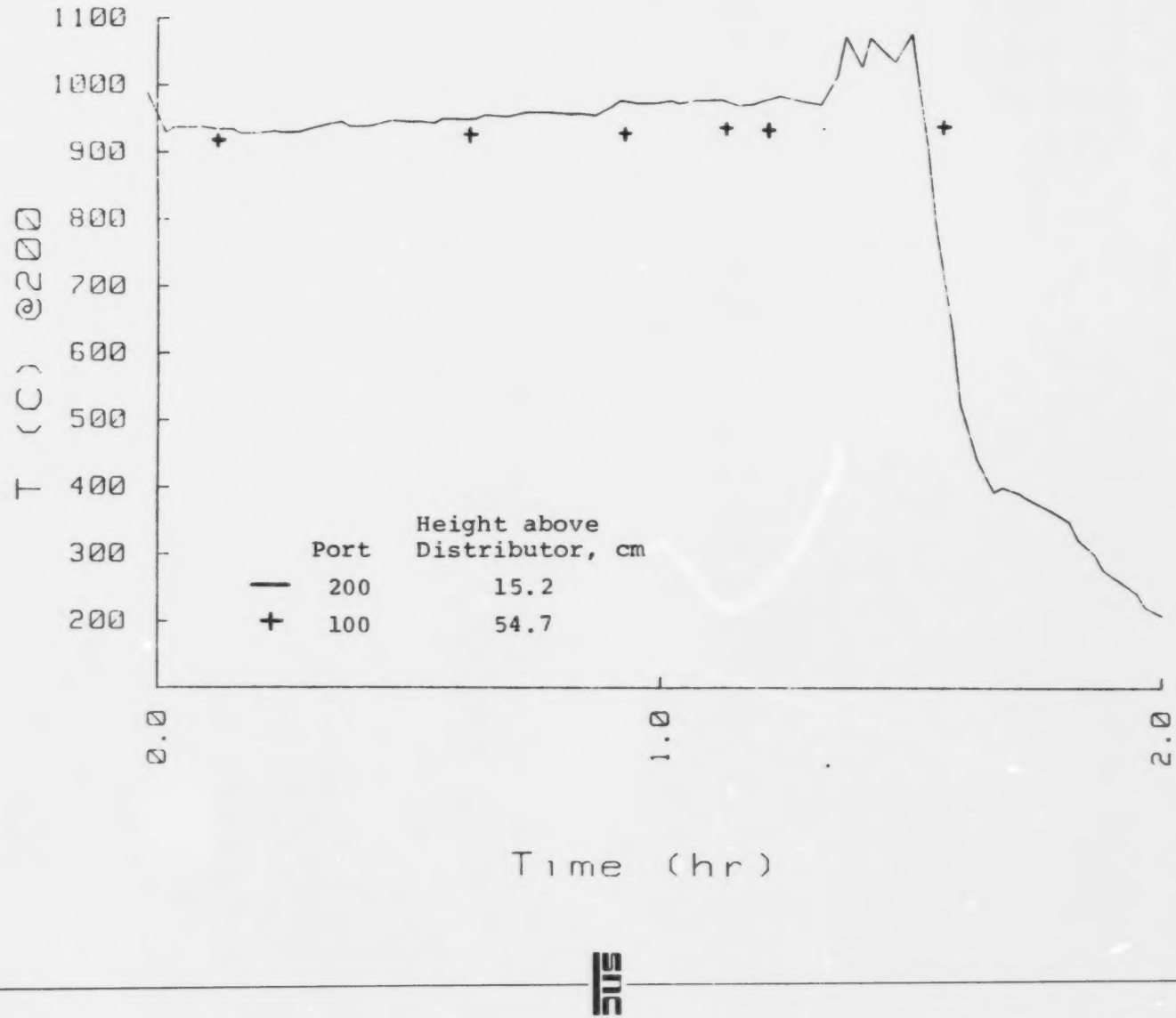


Fig. 5.7 RUN 830614 BED TEMPERATURE (C)



in temperature at the lower level. This apparent decrease was caused by an insulating layer of unfluidized agglomerates around the temperature probe. The probe responded satisfactorily when the insulating layer was removed.

The feed mode was changed from underbed to overbed in an attempt to improve the dispersion of the coal particles in the bed. Run 830705 which was the first run to use this feed mode gave encouraging results. Although there was an early indication of agglomerate formation (Fig. 5.2), the thermocouple readings in the upper levels of the bed remained reasonably steady and the run was continued under gasification conditions for five hours. The run was terminated at this point when the thermocouple at the second level indicated agglomerate formation in the vicinity of the probe. Additional support for an approach to steady conditions is given by the agreement between the stack gas analyses at run times of 1.43 hr and 3.68 hr (Fig. 5.8 and Fig. 5.9). However, this was the only run in which steady operating conditions were approached. In the remaining three runs, in which the gasifier was operated with overbed feed, running times under gasification conditions were not significantly different from the earlier runs with underbed feed.

A contributing factor to consider in an examination of particle agglomeration is the relatively low temperatures characterizing the ash fusibility of Minto coal (Table 4.4). The temperatures given in Table 4.4 for Minto coal apply to oxidizing

Fig. 5.8 Continuous analysis of stack gas.

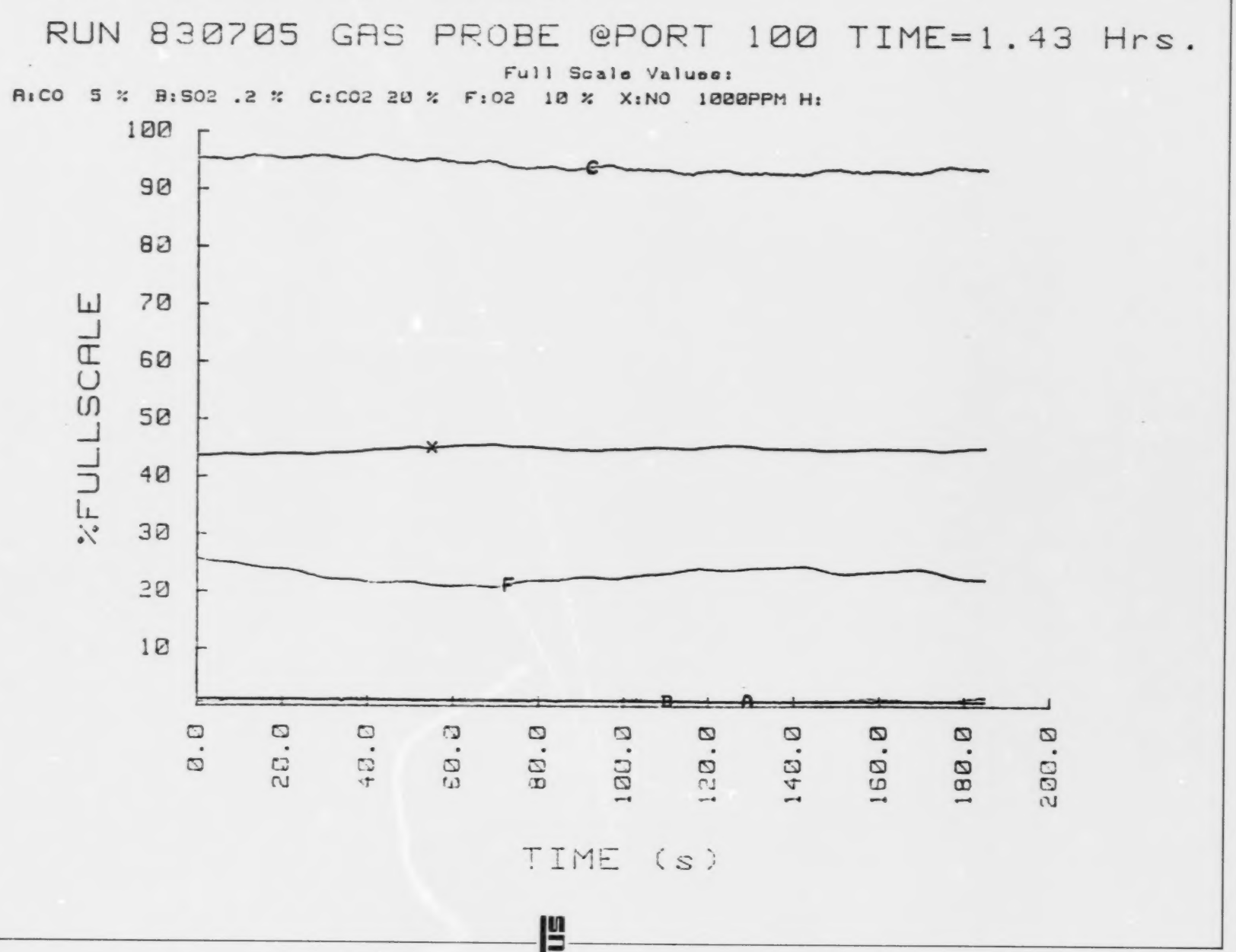
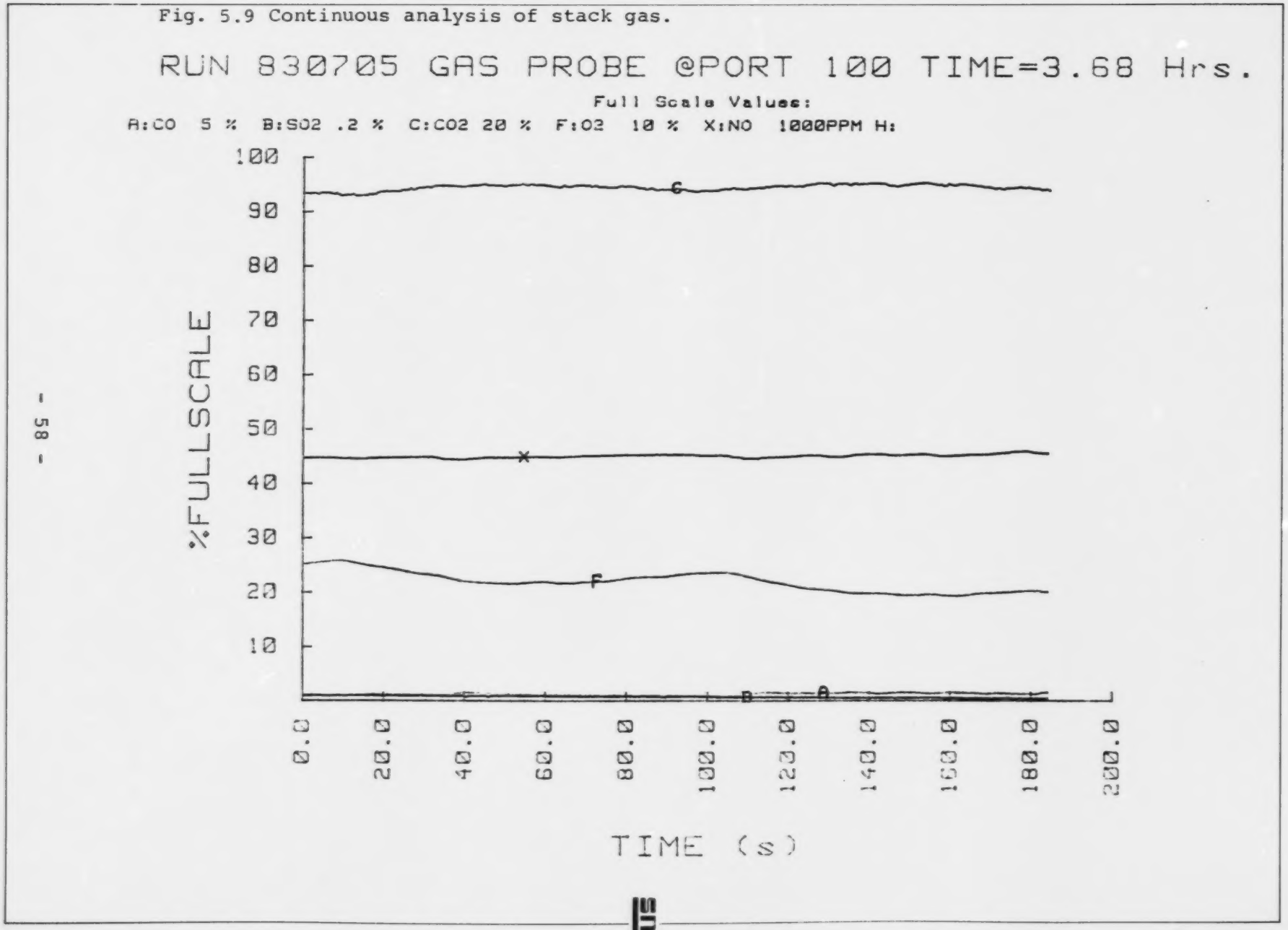


Fig. 5.9 Continuous analysis of stack gas.



conditions. Corresponding temperatures are given for Devco coal under both oxidizing and reducing conditions. The temperatures under reducing conditions are lower by 168 °C for the initial temperature and 178 °C for the spherical softening temperature. If the corresponding temperatures for Minto coal are reduced by these values, estimates of 898 °C and 921 °C are obtained for the initial temperature and the spherical softening temperature.[1] Since the specified bed temperature for the gasification tests was 900 °C, the relatively low ash softening temperature of Minto coal could be a contributing factor in agglomerate formation.

Agglomeration of the coal particles has an adverse effect on the quality of fluidization and mixing in the bed. This results in poor dispersion, the development of local zones of high temperature and subsequent loss of temperature control in the bed. We were unsuccessful in overcoming this problem and the short running times under gasification conditions precluded an approach to steady operating conditions.

SMC

1. After completion of the report information has become available on the ash fusibility of Minto coal under reducing conditions (Anthony 1983). An initial temperature of 1052 °C and a spherical softening temperature of 1071 °C were obtained for a sample with a proximate analysis of 22.5% ash, 28.5% volatile matter and 49.0% fixed carbon (dry basis) and a sulfur content of 7.1% (dry basis).

6. RESULTS AND DISCUSSION

6.1 Synthesis gas

Synthesis gas compositions for the three levels of Ca/S used in the tests are presented in Fig. 6.1 through Fig. 6.3. The superficial fluidization velocity and the steam-to-air ratio for the three runs (830705, 830707, 830714) were constant at 1.6 m/s and 0.38 mol/mol, respectively. Because of the short running times and the unsteady operation of the gasifier, the results should be treated qualitatively. With the exception of H_2, the gas composition is reasonably steady in the later stages of run 830705. One can only speculate on an explanation for the fluctuations in the H_2 concentration. Agglomerate formation and the periodic development of local zones of high temperature could increase the H_2 output from the H_2O - C reaction. The CH_4 concentration is lower than expected from fluid-bed gasification. Although the gasifier was operated at fluidization velocities well above the minimum, agglomerate formation had an adverse effect on the quality of fluidization and mixing in the bed. CO_2 from the calcination of limestone has a diluting effect on the CH_4 concentration (Fig. 6.1). It is evident from Fig. 6.3 that the use of limestone as a sulfur sorbent produces a gas with a significantly lower concentration of H_2S. The different levels of CO_2 in Fig. 6.2 may be attributed to the CO_2 produced by the calcination of the limestone.

6.2 Sulfur capture

A significant portion of the sulfur in the coal feed is

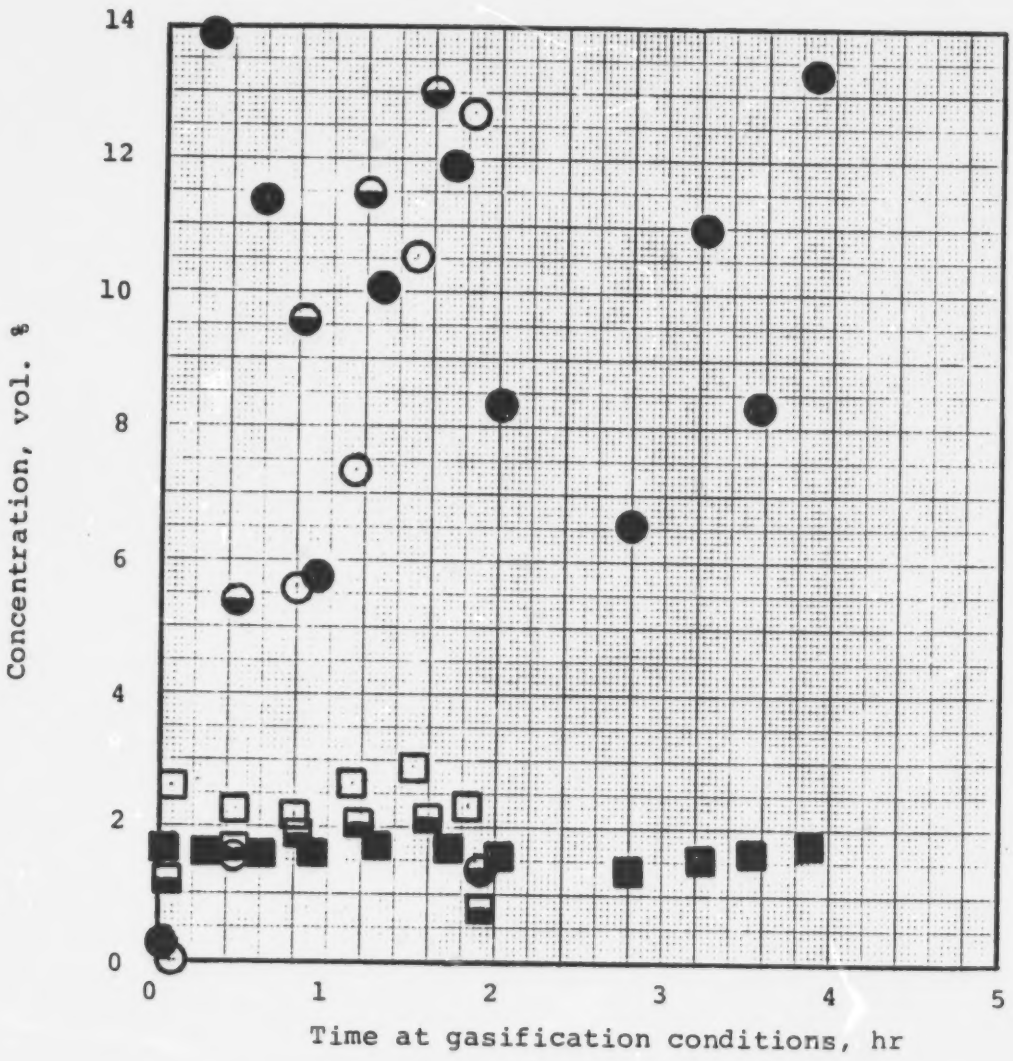


Fig. 6.1. Composition of synthesis gas.

Symbol	Run	Species	Ca/S Ratio
●	830705	H_2	3.0
◐	830707		1.5
○	830714		0.0
■	830705	CH_4	3.0
◧	830707		1.5
□	830714		0.0

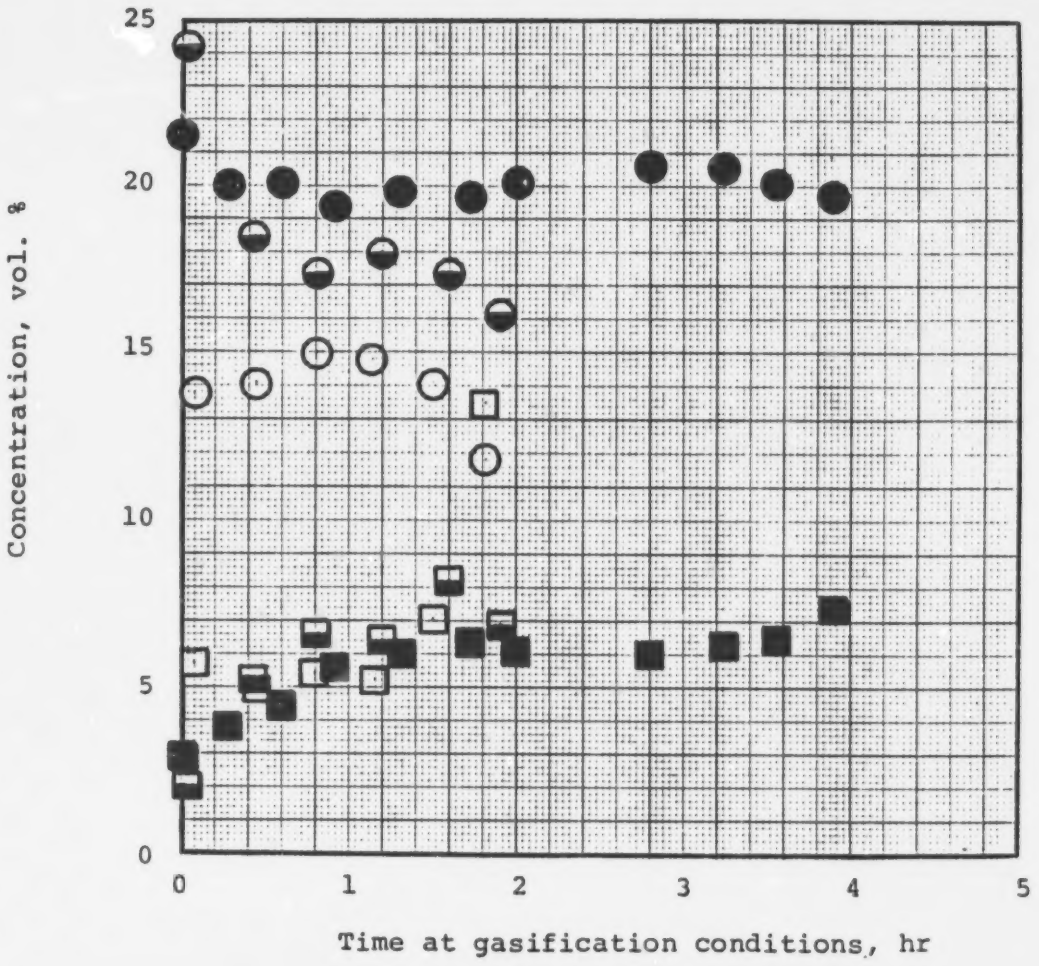


Fig. 6.2. Composition of synthesis gas

Symbol	Run	Species	Ca/S Ratio
●	830705	CO_2	3.0
◒	830707		1.5
○	830714		0.0
■	830705	CO	3.0
◨	830707		1.5
□	830714		0.0

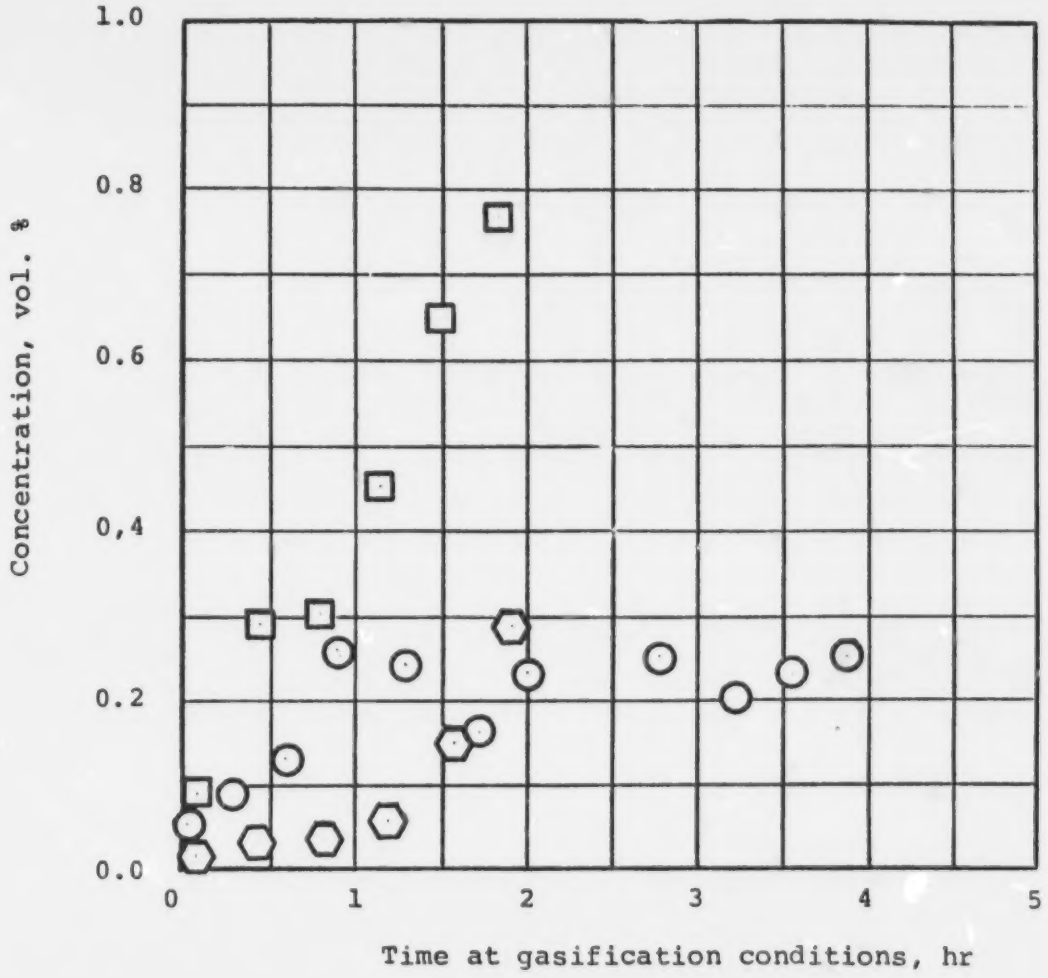


Fig. 6.3. Hydrogen sulfide composition of synthesis gas

Symbol	Run	Ca/S Ratio
○	830705	3.0
⬡	830707	1.5
□	830714	0.0

retained by the bed material. In an attempt to determine the level of sulfur capture and the redistribution of the sulfur in the coal among various sulfur-containing species, approximate sulfur balances were carried out. Steady state, i.e. no accumulation, was assumed and any sulfur leaving in the solids in the synthesis gas stream from the cyclone was neglected. Two calculation schemes were used for estimating the flowrate of dry synthesis gas. In the first calculation (Scheme 1), N_2 was used as a tie substance. The flowrate of synthesis gas was estimated from the known flowrate of N_2 in the incoming air and the N_2 concentration in the synthesis gas from the chromatographic analysis. In the second calculation (Scheme 2), the air flowrate to the bed was adjusted to account for the chemical reactions occurring in the bed. Again, use was made of the chromatographic analysis of the synthesis gas. Any depletion of the O_2 in the incoming air by CO formation was neglected, i.e., it was assumed that all of the CO was formed by the reaction of steam with carbon. The error in this assumption is small since the CO concentration in the synthesis gas is low, of the order of 6 % (Fig. 6.2). CO_2 from the calcination of the limestone was included in the estimation of the flowrate of the synthesis gas. Complete calcination of the carbonates in the limestone was assumed. Estimates of the uncaptured sulfur were obtained from the H_2S leaving in the synthesis gas stream. A third estimate of the sulfur capture was obtained by a sulfur balance on the solids streams: coal, bed overflow and cyclone dust. Sulfur analyses

Table 6.1. Estimates of the sulfur capture, expressed as a percentage of the sulfur in the coal feed

Run 830627:

Time Interval EDT, hr	Feed Ca/S mol/mol	$\frac{S_{Out}}{S_{In}}$	$S_{Cap.}$ (Scheme 1) %	$S_{Cap.}$ (Scheme 2) %	$S_{Cap.}$ (Solids) %
0.10 - 1.26	2.90	--	97.0	96.7	--
1.26 - 1.89	2.71	0.92	98.7	98.6	91.0
1.89 - 2.62	2.86	--	--	--	87.8
2.62 - 3.45	1.58	--	71.8	68.9	--

Run 830705:

Time Interval EDT, hr	Feed Ca/S mol/mol	$\frac{S_{Out}}{S_{In}}$	$S_{Cap.}$ (Scheme 1) %	$S_{Cap.}$ (Scheme 2) %	$S_{Cap.}$ (Solids) %
0.07 - 0.98	3.36	1.08	92.4	91.9	99.7
0.98 - 1.90	3.27	0.93	87.9	87.4	80.6
1.90 - 2.90	3.20	0.87	86.0	85.8	73.1
2.90 - 3.48	3.66	0.93	87.7	87.1	80.6

Run 830707:

Time Interval EDT, hr	Feed Ca/S mol/mol	$\frac{S_{Out}}{S_{In}}$	$S_{Cap.}$ (Scheme 1) %	$S_{Cap.}$ (Scheme 2) %	$S_{Cap.}$ (Solids) %
0.47 - 0.92	1.48	--	98.4	98.3	--
0.92 - 1.42	1.59	--	97.2	97.3	--
1.42 - 1.90	1.86	--	91.3	91.0	--

Run 830714:

Time Interval EDT, hr	Feed Ca/S mol/mol	$\frac{S_{Out}}{S_{In}}$	$S_{Cap.}$ (Scheme 1) %	$S_{Cap.}$ (Scheme 2) %	$S_{Cap.}$ (Solids) %
0.18 - 0.68	0.0	--	89.6	89.6	--
0.68 - 1.18	0.0	--	84.1	84.3	--
1.18 - 1.68	0.0	0.73	67.1	65.8	38.6

Note:

EDT, elapsed decimal time, measured from the switchover to gasification

$\frac{S_{Out}}{S_{In}}$, material balance check, sulfur in the bed overflow plus sulfur in the cyclone dust plus sulfur in the synthesis gas divided by the sulfur in the coal feed

of samples of the bed overflow and cyclone dust are given in Appendix A. The estimates of the sulfur capture, expressed as a percentage of the sulfur in the coal feed, are summarized in Table 6.1.

Analyses for various sulfur-containing species, in the coal (Table 4.3), in the bed overflow and cyclone dust (Appendix A), and in the synthesis gas (Appendix B), were applied to the sulfur balance results to obtain an indication of the redistribution of sulfur in the gasification process. The results for the various sulfur forms, expressed as a percentage of the total sulfur in the coal feed, are presented in Table 6.2. In the analyses the sulfur forms are pyritic sulfur, organic sulfur, $SO_4^=$, $S^=$ and H_2S.

In an examination of the numerical values in Tables 6.1 and 6.2, the approximate nature of the sulfur balances should be borne in mind. The most questionable assumption made was steady-state operation of the gasifier. This is the reason that three calculation schemes were used. Taken together the three schemes made use of the analyses of all the incoming and outgoing streams. The stream flowrates used were measured flowrates with the exception of the flowrate of the synthesis gas stream which was estimated. Qualitatively, the results show that a significant fraction of the sulfur in the coal is captured by the bed material. Of particular interest is the conversion of the sulfur from the pyritic form in the coal to $S^=$ in the bed

Table 6.2. Sulfur forms, expressed as a percentage of the total sulfur in the coal feed

Coal				
Pyr.	Org.	$SO_4^=$	$S^=$	Tot.
52.87	11.78	34.55	0.80	100

Run 830627:

Time Interval	Bed Overflow					Cyclone Dust					Syn. Gas
EDT, hr	Pyr.	Org.	$SO_4^=$	$S^=$	Tot.	Pyr.	Org.	$SO_4^=$	$S^=$	Tot.	H_2S
1.26 - 1.89	0.41	9.75	0.84	38.50	49.48	1.08	10.69	2.96	25.83	40.55	1.41
1.89 - 2.69	0.26	8.70	0.37	33.10	42.42	1.54	26.36	4.80	11.69	44.38	—

Run 830705:

0.070 - 0.98	0.64	14.02	1.29	53.50	69.46	0.31	3.78	1.47	13.51	19.08	8.06
0.98 - 1.90	0.64	8.19	0.11	41.32	50.28	0.48	6.17	1.31	22.30	30.26	12.86
1.90 - 2.90	0.44	19.43	0.34	18.54	38.77	0.57	5.48	1.17	19.31	26.53	14.20
2.90 - 3.48	0.83	11.71	0.26	31.60	43.87	0.56	7.57	0.87	21.96	30.96	12.87

Run 830714:

1.18 - 1.68	1.31	13.31	0.12	12.13	26.86	0.71	4.81	0.87	5.52	11.92	34.23

overflow and cyclone dust. If the Fe retains the same oxidation state as in pyrites, it is reasonable to postulate that some of the $S^{=}$ in the bed overflow and cyclone dust is associated with Ca. Some measure of the Fe associated with $S^{=}$ in the bed may be obtained from the results for run 830714. In this run no limestone was fed to the bed and the initial charge was a silica sand.

6.3 Agglomerate formation

Agglomerate formation and the consequential operating problems have been recurring themes throughout this report. It is evident from our lack of success in overcoming these problems that process modifications are required for the gasification of Minto coal. Forney et al. (1967) report on the fluid-bed gasification at 870 °C of a caking coal (Pittsburgh seam coal, Bruceton mine) with a free swelling index (FSI) of 8. They found that it was necessary to stage the process and pretreat the coal to remove the caking properties for satisfactory operation of the gasifier. Two methods of pretreatment in a fluid bed were used: 1) fluidization of the coal in steam plus air or oxygen at 430 °C for 5-10 min, and 2) fluidization of the coal in air at 240 °C for 30-40 min. It was necessary to reduce the FSI of the coal to approximately 1 in order to prevent agglomeration in the gasifier. Pretreated coals with FSI's of 2.0 and 1.5 caked in the gasifier. Thus, the difficulties which we encountered with Minto coal with an FSI of 5 to 5.5 (Tibbetts and Montgomery, 1976) are not unexpected.

7. CONCLUSIONS AND RECOMMENDATIONS

A number of conclusions may be drawn from the results of this investigation.

1. Particle agglomeration is a major problem in the gasification of Minto coal. It leads to defluidization, the development of local zones of overheating, and finally loss of temperature control in the bed. Minto coal is a moderately strongly caking coal (free swelling index 5.0 - 5.5) and has a low ash fusion point (initial temperature 1052 °C under reducing conditions). Such coals commonly exhibit strong agglomeration tendencies and Minto coal turned out to be no exception.

2. The use of limestone as a sorbent effectively reduces the H_2S content of the synthesis gas. At a Ca/S ratio of 3 the sulfur capture by the bed solids is of the order of 90%. Based on the dry analysis of the synthesis gas for run 830705 (Appendix B, run time 3.88 hr) the higher or gross heating value was estimated at 3 230 kJ/m³ (86.7 Btu/ft³) or 2 780 kJ/kg (1 200 Btu/lb); reactants and products at 15.6 °C (60 °F) and 101.3 kPa (1 atm).

3. It appears that at least 3 hours of operating time under gasification conditions are required for the attainment of steady-state conditions. An additional 3 to 5 hours were required to reach gasification conditions from a cold start.

Modifications to the process are required for the gasification of Minto coal. The following recommendations are made for consideration in further gasification studies.

1. One modification which is suggested is pretreatment of the coal in a fluid bed to remove caking properties, e.g., Forney et al. (1967). In the modification the volatiles would be retained in the system by injecting the product gas from the pretreater into the gasifier.

2. A more vigorous means of introducing and dispersing the coal is required to counteract the agglomeration tendencies. Rapid dispersion can be promoted by pneumatically feeding the coal in a upwardly directed jet from the bottom of the bed. The jet would form a spout and the bed would then operate as a spouted-fluid bed. The separation of the coal particles can be promoted by deepening the bed, thereby increasing the holdup of inert solids and lowering the concentration of coal particles.

3. Gasification in a circulating bed is another mode of operation which should be given consideration.

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APPENDIX A

Analysis of Selected Samples of Bed Solids and Cyclone Dust

Table A.1. Analyses of samples of bed overflow (ash) and cyclone dust (cyc)

Sample	C, %	H, %	Sulfur, %				
			Pyr.	Org.	$SO_4^=$	$S^=$	Total
830627-ash 0030-0106	--	--	0.07	1.64	0.14	6.48	8.33
830627-ash 0131-0225	--	--	0.05	1.65	0.07	6.28	8.05
830627-cyc 0048-0126	--	--	0.16	1.59	0.44	3.84	6.03
830627-cyc 0126-0210	--	--	0.17	2.91	0.53	1.29	4.90
830705-ash 0015-0045	10.63	0.57	0.07	0.89	1.33	4.13	6.42
830705-ash 0110-0145	14.11	0.31	0.08	1.74	0.16	6.64	8.62
830705-ash 0153-0233	23.05	0.43	0.12	1.53	0.02	7.70	9.37
830705-ash 0258-0328	23.31	0.58	0.10	4.43	0.08	4.23	8.84
830705-ash 0400-0430	22.28	0.58	0.16	2.27	0.05	6.13	8.51
830705-cyc 0010-0040	21.67	0.32	0.12	0.91	0.89	2.04	3.96
830705-cyc 0040-0135	22.03	0.35	0.09	1.08	0.42	3.86	5.45
830705-cyc 0135-0230	20.52	0.37	0.10	1.27	0.27	4.59	6.23
830705-cyc 0230-0330	19.70	0.36	0.15	1.41	0.30	4.97	6.83
830705-cyc 0330-0405	17.77	0.34	0.13	1.76	0.20	5.10	7.19
830714-ash 0000-2330	--	--	0.21	2.14	0.02	1.95	4.32
830714-cyc 2222-2345	--	--	0.22	1.48	0.27	1.70	3.67

Analysis by analyst A

APPENDIX B

Analyses of Synthesis Gas Stream

SYNTHESIS GAS CONCENTRATIONS **RUN** 830606

GAS CHROMATOGRAPH

CLOCK TIME	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
17:58		53.14	1.25	4.24	15.08	0		
18:50		60.74	1.55	7.93	17.45	0.008		
19:78		61.87	1.73	9.33	17.24	0.020		

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{DM}$			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS

RUN 830608

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	H_2O_m
00:07	8.54	65.27	1.33	3.07	18.82	0	2.97	1.83
00:44	8.74	63.66	1.23	6.92	13.04	0	6.41	4.04
01:24	8.85	56.92	1.68	5.93	17.66	0	8.96	8.83
02:08	8.67	51.32	1.19	5.56	19.00	0	14.04	
	H_2O_m estimated from tailing peak							

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	H_2O_{SY}	H_2O_{MN}			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS **RUN** 830614

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
0.38	1.94	53.31	1.67	7.65	12.93		22.50	
1.08	3.97	52.49	1.81	8.90	12.86		19.97	
1.80	2.08	53.32	0.63	11.72	11.01		21.24	

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{S/}$	$H2O_{SN}$			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS

RUN 830621

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	H_2O_m
00.10	11.55	55.68	2.12	3.51	17.31	0	9.83	
00.70	13.01	51.30	1.20	6.22	14.23	0.014	14.03	
01.32	11.67	54.16	0.32	8.62	11.84	0.060	13.40	

H_2O_m from comparison with dry CO_2 concentration

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{sy}$	$H2O_{cm}$			
00.53	18.06	0.12	1.62	7.05 (error)	2.08			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	
1.03	0.4115	0.0366	15.81	0.026	0.028	1.39	0.096	

SYNTHESIS GAS CONCENTRATIONS

RUN 830623

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	H_2O_m
00.03		48.63	1.30	1.44	15.66			
00.53	3.29	52.24	1.61	3.61	15.34	0.039		
01.12	12.48	49.05	1.57	4.83	13.94	0.120	18.01	21.7
01.38	9.11	51.21	1.34	5.97	13.90	0.160	18.31	21.7

* H_2O_m calculated by comparing $[CO_2]$ in dry gas with $[CO_2]$ in wet gas

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{sy}$	$H2O_{cm}$			
00.12	20.46	0.12	1.77				*CO_2	over
01.14	17.77			34.99	2.71			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	
1.78	0.0421	0.0371	17.30			4.23	0.081	

SYNTHESIS GAS CONCENTRATIONS **RUN** 830627

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
0.13		42.43	1.00	1.40	14.00	0.065		
0.70	10.66	35.81	0.87	2.22	10.58	0.028	39.83	
1.50	5.45	30.52	0.80	2.84	8.68	0.014	51.70	
2.83	9.80	37.01	1.00	3.98	10.85	0.285	37.08	

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{CVO}$			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS **RUN** 830705

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
00.02	0.21	66.73	1.66	2.92	21.55	0.053		
00.28	13.88	64.43	1.64	3.82	20.09	0.089		
00.60	11.40	64.23	1.61	4.49	20.11	0.13		
00.92	5.78	60.74	1.64	5.58	19.40	0.26	Dry	gas
01.30	10.09	60.10	1.73	6.05	19.87	0.24		
01.73	11.89	58.83	1.73	6.43	19.71	0.16		
02.01	8.37	58.90	1.60	6.17	20.20	0.23		
02.78	6.30	60.68	1.46	6.03	20.72	0.25		
03.23	11.00	59.65	1.57	6.31	20.65	0.20		
03.56	8.36	58.75	1.66	6.47	20.16	0.23		
03.88	13.33	56.93	1.73	7.40	19.67	0.25		

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{NO}$		CO	
00.03							2.82	
02.40				12.7				

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	
1.43	0.0584	0.0016	18.83	447	453	2.24		
3.68	0.0633	0.0010	18.83	450	226	2.10		

SYNTHESIS GAS CONCENTRATIONS **RUN** 830707

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
00.05		65.87	1.22	2.06	24.29	0.013		
00.43	5.39	63.42	1.65	5.28	18.48	0.031		
00.82	9.57	59.65	1.87	6.68	17.43	0.034		
01.18	11.47	57.65	2.06	6.45	18.02	0.054		
01.58	13.07	54.83	2.14	8.22	17.36	0.149		
01.91	1.39	66.75	0.90	6.96	16.15	0.285		

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{SW}$	CO		
00.08						2.38		
01.20	18.94							

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS **RUN** 830712

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
00.58		66.49	1.16	1.85	25.40			
00.98		61.72	1.52	3.87	21.62	0.003		
01.45	1.709	60.22	1.75	7.47	18.44	0.055		
01.85	4.38	56.43	1.89	8.16	18.60	0.008		
02.30	3.41	58.67	1.91	6.93	20.29	0.012		

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{SM}$			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	

SYNTHESIS GAS CONCENTRATIONS **RUN** 830714

GAS CHROMATOGRAPH

Run Time	H2	N2	CH4	CO	CO2	H2S	ΔH2O	$H2O_m$
00.08	0.01	71.53	2.61	5.70	13.77	0.089		
00.45	1.58	66.03	2.29	5.13	14.08	0.288		
00.80	5.62	63.37	2.19	5.44	14.99	0.301		
01.15	7.34	62.43	2.63	5.25	14.78	0.426		
01.50	10.55	58.43	2.91	7.05	14.11	0.648		
01.83	12.71	53.08	2.34	13.48	11.85	0.766		

CONTINUOUS MONITORING

Run Time	CO2	SO2	HC	$H2O_{SY}$	$H2O_{NR}$			

INCINERATOR GAS CONCENTRATIONS

Run Time	CO	SO2	CO2	NO	NOX	O2	HC	
00.90	0.53	0.420	16.02	286	332	2.89		
01.60	2.12	0.393	15.40		267	3.79		
01.95	1.60	0.376	14.66	264		0.82		

APPENDIX C

Proximate Analyses of Bed Solids and Cyclone Ash

FBC SOLID ANALYSIS RUN: 830606

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	17:30-18:20	0	88.75	11.25	3.96			
	σ		0.17	0.17				
ASH	18:20-18:47	0	88.86	11.14	4.34			
ASH	18:47-19:45	0	88.54	11.46	3.56			
CYC	17:35-18:01							
CYC	18:01-18:24							
CYC	18:24-18:41	0	84.67	15.33	5.19			
			0.06	0.06				
CYC	18:41-18:59	0	82.83	17.17	5.60			
CYC	18:59-19:20	0	81.34	18.65	5.29			
CYC	19:20-19:40	0	82.68	17.32	4.24			

COMMENTS:

σ is the variance of the sample determinations.

FBC SOLID ANALYSIS RUN: 830608

Exit Streams Part A (Proximate – Organic)

	Sample		(% by weight on a moisture free basis)						
	Type	No.	H2O	ASH	F.C.	S	C	H	N
	ASH	12:49-13:05	0	86.34	13.66				
	ASH	13:05-13:25							
	ASH	13:25-13:53	0	82.14	17.86	5.03			
		σ		0.245					
	ASH	13:53-14:30	0	77.26	22.74	4.95			
	ASH	14:30-14:35	0	82.61	17.39				
x	ASH	14:35-15:11	0	79.92	20.08	4.05			
		σ		0.514		0.12			
	ASH	15:11-16:14	0	94.71	5.29				
		σ		0.085					

COMMENTS:

FBC SOLID ANALYSIS RUN: 830608

Exit Streams Part A (Proximate – Organic)

	Sample		(% by weight on a moisture free basis)						
	Type	No.	H2O	ASH	F.C.	S	C	H	N
	CYC	13:10-13:29	0	78.00	22.00				
	CYC	13:29-13:50	0	78.15	21.85				
	CYC	13:50-14:15	0	75.39	24.61	3.30			
	CYC	14:15-14:45	0	76.02	23.98	3.62			
x	CYC	14:45-15:05	0	79.62	20.38	3.82			
		σ				0.010			
	CYC	15:05-15:20	0	87.67	12.33				
	CYC	15:20-15:42	0	85.94	14.06				
	CYC	15:42-16:05	0	89.32	10.68				
		σ		0.075					

COMMENTS: _____

FBC SOLID ANALYSIS RUN: 830610

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	11:00-11:20	0	91.60	8.40				
	σ		0.586					
CYC	11:00-11:20	0	81.45	18.55				
								830613
ASH	00:18-00:59	0	84.93	15.07				
CYC	23:03-00:35	0	83.59	16.41				

COMMENTS:

FBC SOLID ANALYSIS RUN: 830614

Exit Streams Part A (Proximate – Organic)

	Sample		(% by weight on a moisture free basis)						
	Type	No.	H2O	ASH	F.C.	S	C	H	N
	ASH	23:00-23:15	0	91.38	8.62	3.51			
		σ		0.167		0.24			
	ASH	23:15-23:50	0	100	0	8.96			
		σ		0					
x	ASH	23:50-00:15	0	99.22	0.78	10.8			
		σ		0.545					
	ASH	00:15-01:05	0	92.50	7.50	10.8			
		σ				0.57			
	CYC	23:00-23:30	0.34	65.85	34.15	3.49			
x	CYC	23:30-24:00	0.22	70.05	29.95	4.56			
		σ		0.917					

COMMENTS: _____

FBC SOLID ANALYSIS RUN: 830621

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	21:30-22:00							
ASH	22:00-22:30							
ASH	22:30-22:58	0	86.38	13.62	2.90			
	σ				0.20			
CYC	21:51-22:20	0	67.12	32.88	11.6			
	σ				0.83			

COMMENTS:

FBC SOLID ANALYSIS RUN: 830623

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	22:40-23:10	0	96.69	3.31	7.14			
	σ		0.190		0.13			
ASH	23:10-23:40	0	95.84	4.16	8.26			
	σ		0.596		0.72			
ASH	23:40-00:10	0	96.26	3.74	10.6			
	σ		0.954		0.64			
ASH	00:10-00:40	0	87.56	12.44	11.4			
	σ				0.52			
CYC	23:00-23:30	0	95.94	34.06	4.17			
	σ				0.082			
CYC	23:30-24:00	0.06	58.76	41.24	4.72			
	σ				0.049			
CYC	00:00-00:29	0	69.16	30.84	4.73			
	σ				0.19			

COMMENTS:

FBC SOLID ANALYSIS RUN: 830627

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	23:53-00:30	0	96.89	3.11	6.91			
	σ				0.10			
ASH	00:30-01:06	0	89.53	10.47	8.10			
	σ				0.24			
ASH	01:06-01:31	0	93.29	6.71	8.77			
	σ		0.710		0.14			
ASH	01:31-02:25	0	90.32	9.68	7.34			
	σ				0.22			
ASH	02:25-02:57	0	96.09	3.91	7.81			
	σ				0.17			
CYC	23:38-00:48	0	73.38	26.62	4.60			
	σ		0.373		0.011			
CYC	00:48-01:26	0	78.56	21.44	5.46			
	σ		0.402		0.36			
CYC	01:26-02:10	0.06	71.59	28.41	4.25			
	σ		0.648		0.045			
CYC	02:10-03:00	0	76.16	23.84	5.00			
	σ		0.582		0.11			

COMMENTS:

FBC SOLID ANALYSIS RUN: 830705

Exit Streams Part A (Proximate -- Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	00:15-00:45	0	91.88	8.12	5.28			
	σ				0.37			
ASH	00:45-01:10	0	87.64	12.36	6.02			
ASH	01:10-01:45	0	93.88	6.12	7.21			
	σ				0.25			
ASH	01:45-01:53							
ASH	01:53-02:33	0	90.76	9.24	7.99			
	σ				0.20			
ASH	02:33-02:58	0	87.68	12.32	8.96			
	σ				0.44			
ASH	02:58-03:28	0	86.03	13.97	8.02			
	σ				0.36			
ASH	03:28-04:00	0	81.77	18.23	6.71			
	σ				0.31			
ASH	04:00-04:30	0	84.88	15.12	7.75			
	σ				0.73			
ASH	04:30-04:56	0	84.79	15.21	8.29			
	σ		0.464		0.34			

COMMENTS:

FBC SOLID ANALYSIS RUN: 830705

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
CYC	00:10-00:40	0.16	68.96	31.04	4.04			
	σ				1.25			
CYC	00:40-01:35	0	75.05	24.95	4.39			
CYC	01:35-02:30	0	79.48	20.52	5.58			
	σ				0.77			
CYC	02:30-03:30	0	83.51	16.49	6.25			
	σ				0.35			
CYC	03:30-04:05	0	83.94	16.06	6.05			
	σ		0.386		0.01			

COMMENTS: _____

FBC SOLID ANALYSIS RUN: 830707

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	23:03-23:30	0	97.80	2.20				
ASH	23:30-00:00	0	87.90	12.10				
ASH	00:00-00:29	0	70.57	29.43				
CYC	22:33-23:30							

COMMENTS:

FBC SOLID ANALYSIS RUN: 830712

Exit Streams Part A (Proximate – Organic)

Sample			(% by weight on a moisture free basis)						
Type	No.		H2O	ASH	F.C.	S	C	H	N
ASH	18:45-19:15		0	87.48	12.52	4.73			
ASH	19:15-19:45		0	96.57	3.43	5.67			
ASH	19:45-20:13		0	92.02	7.98	5.79			
	σ			0.350					
ASH	20:13-20:40		0	79.39	20.61	5.27			
ASH	20:40-20:58		0	83.10	16.90	6.19			
CYC	18:45-19:30		0.05	76.36	23.64	4.10			
CYC	19:30-20:28		0	72.02	27.98	3.91			
CYC	20:28-21:28								

COMMENTS:

FBC SOLID ANALYSIS RUN: 830714

Exit Streams Part A (Proximate – Organic)

Sample		(% by weight on a moisture free basis)						
Type	No.	H2O	ASH	F.C.	S	C	H	N
ASH	22:30-23:00	0	84.48	15.52	2.00			
	σ				0.096			
ASH	23:00-23:30	0.07	77.66	22.34	2.70			
ASH	23:30-00:00	0.15	54.90	45.10	4.25			
	σ		0.818		0.24			
ASH	00:00-00:30	0.17	46.01	53.99	4.91			
CYC	22:22-23:45	0.25	67.94	32.06	3.25			
	σ		2.62		0.028			

COMMENTS:

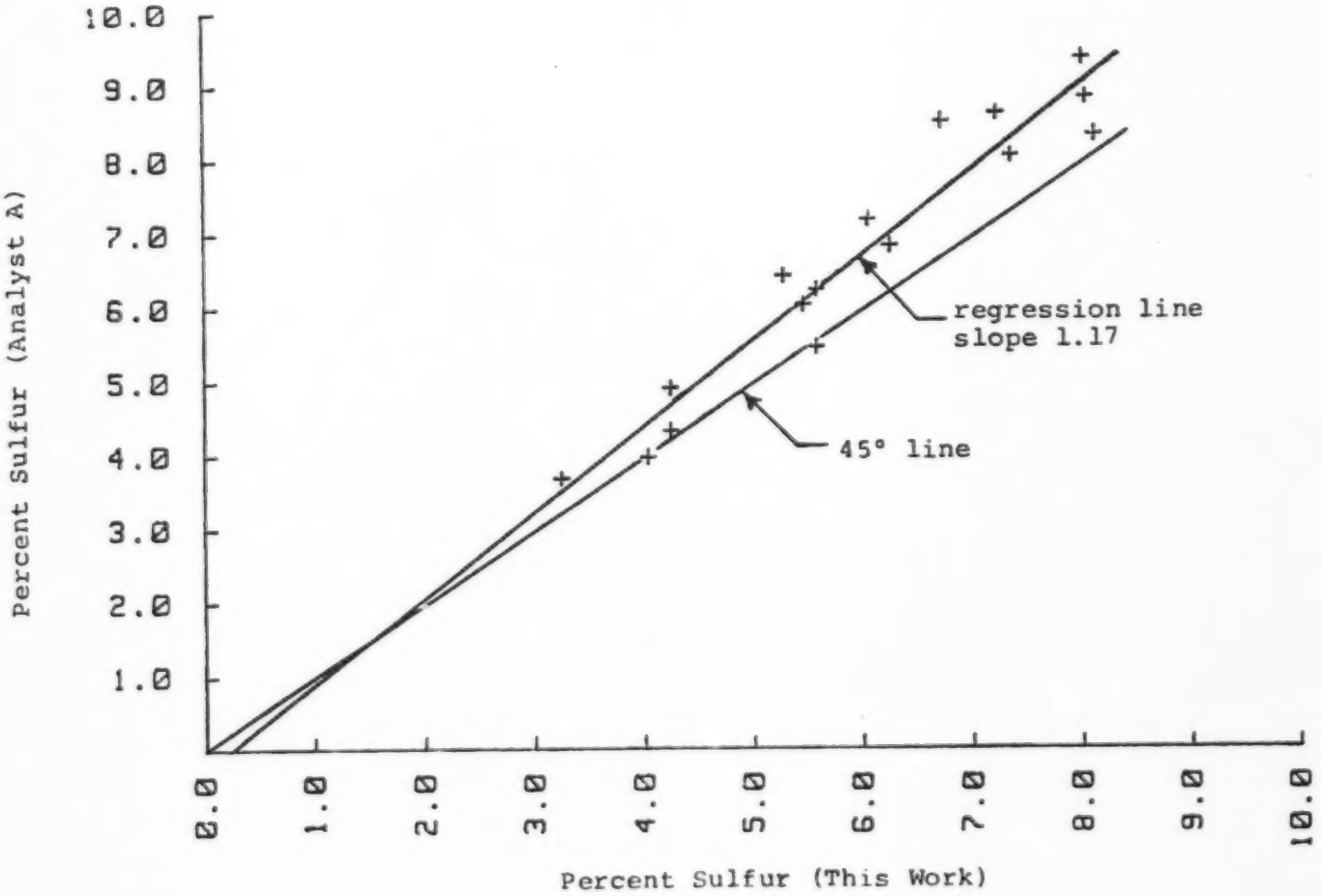


Fig. C.1 Comparison of total sulfur analyses of samples of solids

APPENDIX D

Material Balance Results

Run No.	Duration EDT	Interval EDT	Feed Rates, kg/hr M_C	M_C (maf)	M_A	M_S	M_L
830606	2.6	1.2 - 2.6	112.2	93.1	398	21.5	92.4
830608	2.0	0.6 - 1.7	113.8	94.5	334	16.4	118.4
830614	2.1	0.8 - 2.1	63.8	52.9	205	23.9	25.5
830621	1.8	1.0 - 1.5	74.8	62.1	207	38.5	15.7
830623	1.9	0.5 - 1.5	95.4	79.2	215	43.6	16.3
830627	3.5	0.1 - 1.3	66.9	55.6	194	46.2	38.6
		1.3 - 1.9	68.6	56.9	195	46.4	37.0
		1.9 - 2.6	56.0	46.5	198	47.1	31.8
		2.6 - 3.5	72.0	59.8	197	46.9	22.7
830705	4.3	0.1 - 1.0	67.1	55.7	196	46.7	44.8
		1.0 - 1.9	69.1	57.3	196	46.7	45.0
		1.9 - 2.9	71.0	58.9	195	46.4	45.1
		2.9 - 3.5	67.7	56.2	195	46.4	49.1
830707	2.0	0.5 - 0.9	84.0	69.7	200	47.6	24.6
		0.9 - 1.4	85.7	71.1	201	47.8	27.1
		1.4 - 1.9	104.2	86.5	197	46.9	38.4
830712	2.3	0.5 - 1.5	96.4	80.0	281	15.2	102.0
830714	2.0	0.2 - 0.7	67.9	56.4	194	46.2	0
		0.7 - 1.2	90.0	74.7	194	46.2	0
		1.2 - 1.7	93.6	77.7	194	46.2	0

EDT, elapsed decimal time, hr

M_C, coal

$M_{C(maf)}$, coal, moisture and ash free basis

M_A, air

M_S, steam

M_L, limestone

Run No.	$\frac{M_A}{M_C}$ (maf)	$\frac{M_S}{M_C}$ (maf)	$\frac{M_L}{M_C}$ (maf)	Temp. °C	Sulfur Capture, % (1)	(2)	Carbon Conversion % (3)
830606	4.3	0.23	0.99	950	99	71	27.4
830608	3.5	0.17	1.25	870	100	52	17.7
830614	3.9	0.45	0.48	970	100	30	24.2
830621	3.3	0.62	0.25	930	98	82	15.6
830623	2.7	0.55	0.21	870	91.6	83.5	11.3
830627	3.5	0.83	0.69	845	97.0	97.4	11.1
	3.4	0.82	0.65	920	98.7	91.0	14.2
	4.3	1.01	0.68	820	--	87.8	--
	3.3	0.79	0.38	890	71.8	80.0	15.4
830705	3.5	0.84	0.81	830	92.4	99.7	11.1
	3.4	0.81	0.78	860	87.9	80.6	15.8
	3.3	0.79	0.77	890	86.0	73.1	14.6
	3.5	0.83	0.87	870	87.7	80.6	15.9
830707	2.9	0.68	0.35	895	98.4	--	14.0
	2.8	0.67	0.38	870	97.2	--	14.3
	2.3	0.54	0.44	840	91.3	--	11.9
830712	3.5	0.19	1.28	870	98	74	12.4
830714	3.4	0.82	0	800	89.6	37.8	14.1
	2.8	0.62	0	880	84.1	33.4	11.2
	2.5	0.59	0	895	67.1	38.6	19.6

(1) Sulfur capture from sulfur balance on coal feed and product gas stream; flowrate of product gas estimated by Scheme 1 (p. 64).

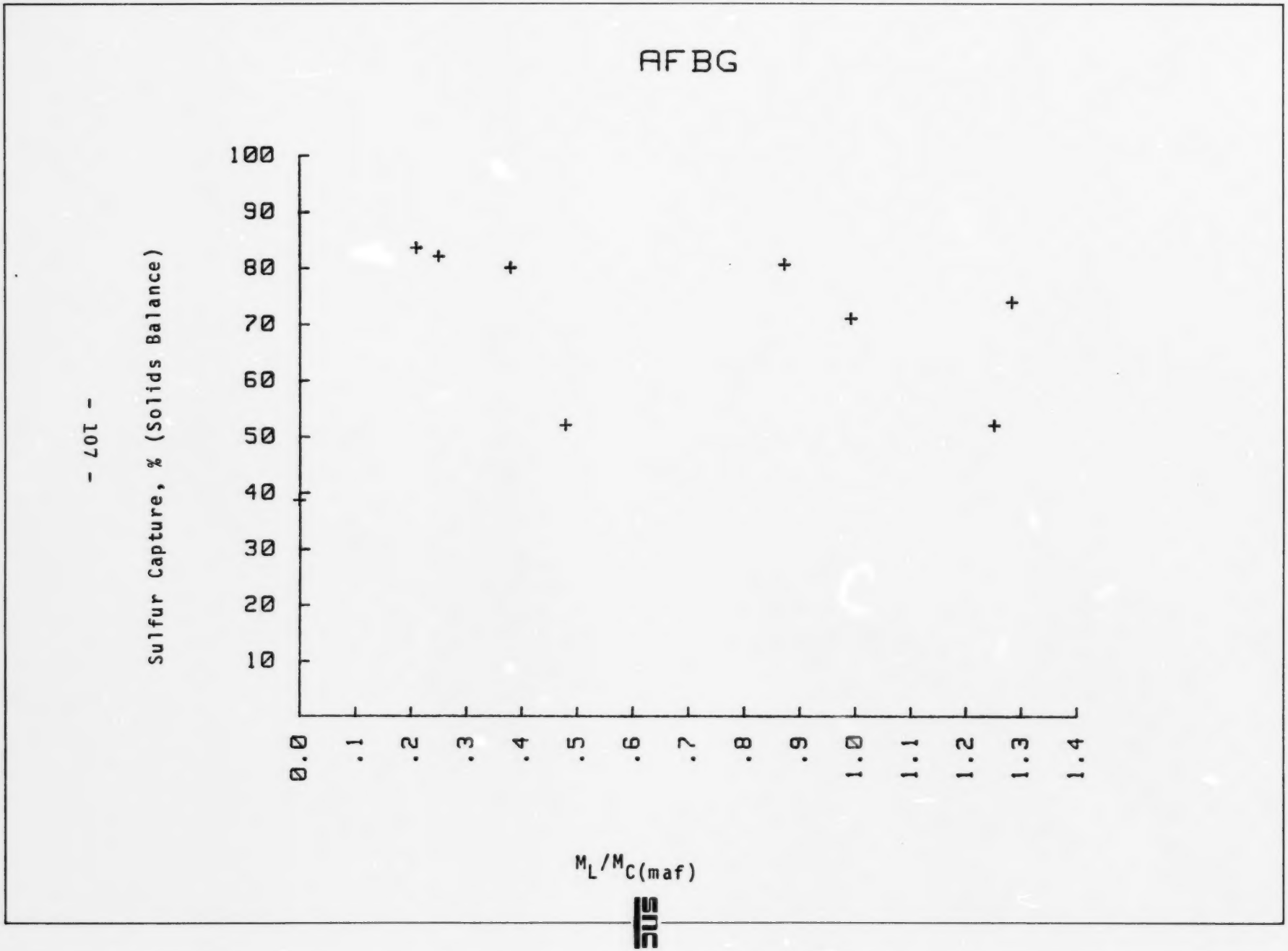
(2) Sulfur capture from sulfur balance on solids streams (coal, bed overflow and cyclone dust).

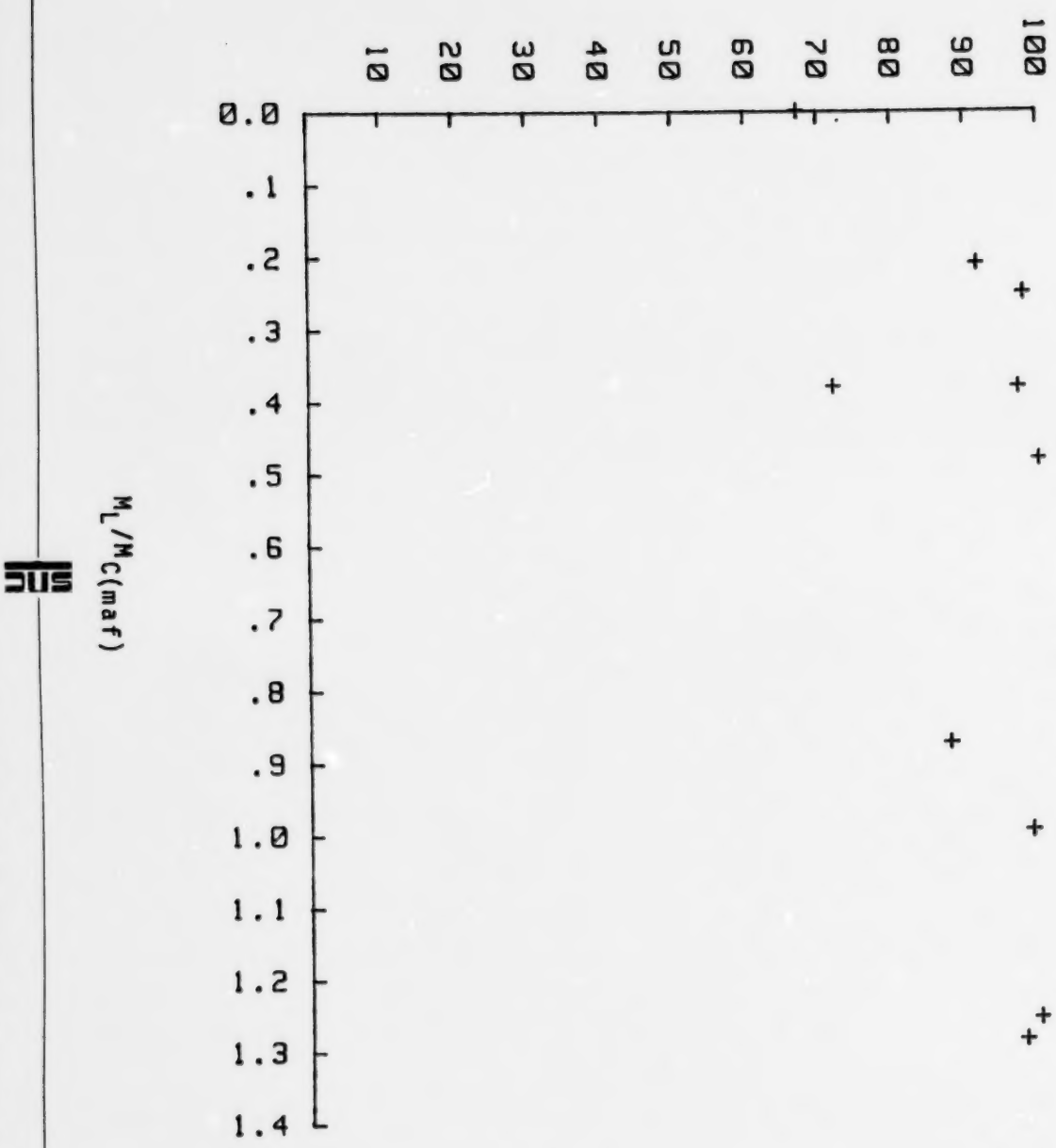
(3) Carbon conversion based on CH_4 and CO content of the product gas stream.

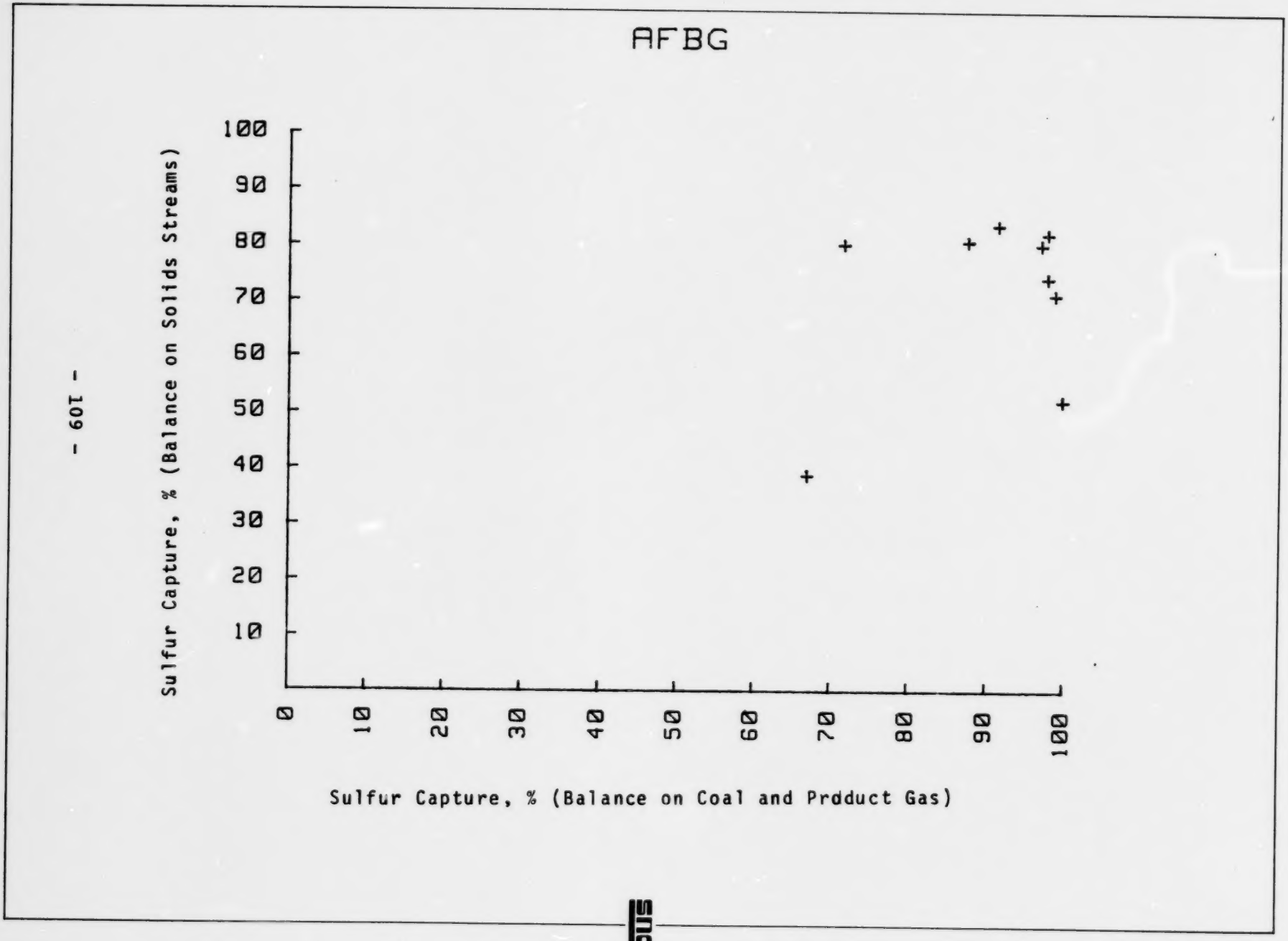
Run No.	Gas Analysis, vol %						Gas Heating Value MJ/m³	Gas Yield m³/kg (maf)
	H_2	N_2	CH_4	CO	CO_2	H_2S		
830606	0	68.94	1.85	9.70	19.5	0.016	1.87	3.85
830608	9.72	62.52	1.85	6.51	19.4	0	2.66	3.42
830614	3.80	66.70	1.53	13.0	15.3	0	2.60	3.58
830621	13.53	62.77	0.88	8.60	15.1	0.040	3.01	3.21
830623	13.19	61.26	1.96	6.60	17.0	0.175	3.17	2.67
830627	17.72	59.52	1.45	3.69	17.6	0.047	3.15	3.54
	11.28	63.19	1.66	5.88	18.0	0.023	2.70	3.28
	-	-	-	-	-	-	-	-
	15.58	58.82	1.59	6.33	17.2	0.474	3.36	3.34
830705	7.82	64.03	1.64	4.20	20.3	0.13	2.10	3.33
	10.99	59.47	1.73	6.24	19.8	0.20	2.78	3.44
	7.34	59.79	1.53	6.10	20.5	0.24	2.26	3.30
	11.00	59.65	1.57	6.31	20.7	0.20	2.73	3.47
830707	9.57	59.65	1.87	6.68	17.4	0.033	2.67	2.82
	11.47	57.65	2.06	6.45	18.0	0.054	2.95	2.90
	7.23	60.79	1.52	7.59	16.8	0.22	2.41	2.2
830712	0.8	60.97	1.64	5.67	20.0	0.29	1.47	3.46
830714	0.80	68.78	2.45	5.42	13.9	0.19	1.72	3.10
	6.48	62.90	2.41	5.35	14.9	0.36	2.42	2.50
	11.63	55.76	2.63	10.27	13.0	0.71	3.80	2.63

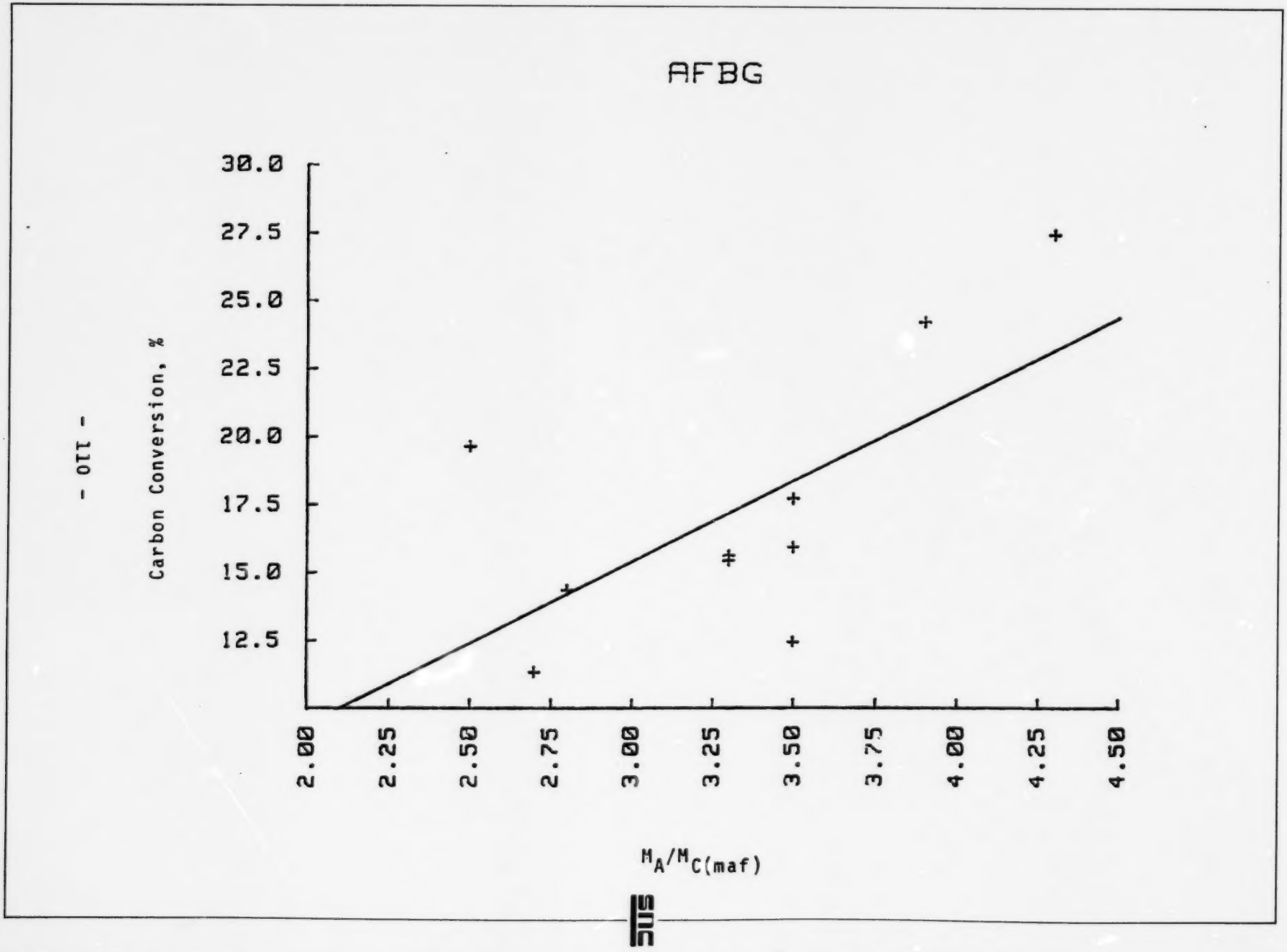
The heating value of the product gas is the higher or gross heating value; reactants and products at 15.6 °C and 101.3 kPa.

Gas yield, m³ at 15.6 °C and 101.3 kPa per kg of coal, moisture and ash free.


AFBG
Sulfur Capture, % (Solids Balance)
100
90
80
70
60
50
40
30
20
10
0.0
.1
.2
.3
.4
.5
.6
.7
.8
.9
1.0
1.1
1.2
1.3
1.4
$M_L/M_{C(maf)}$

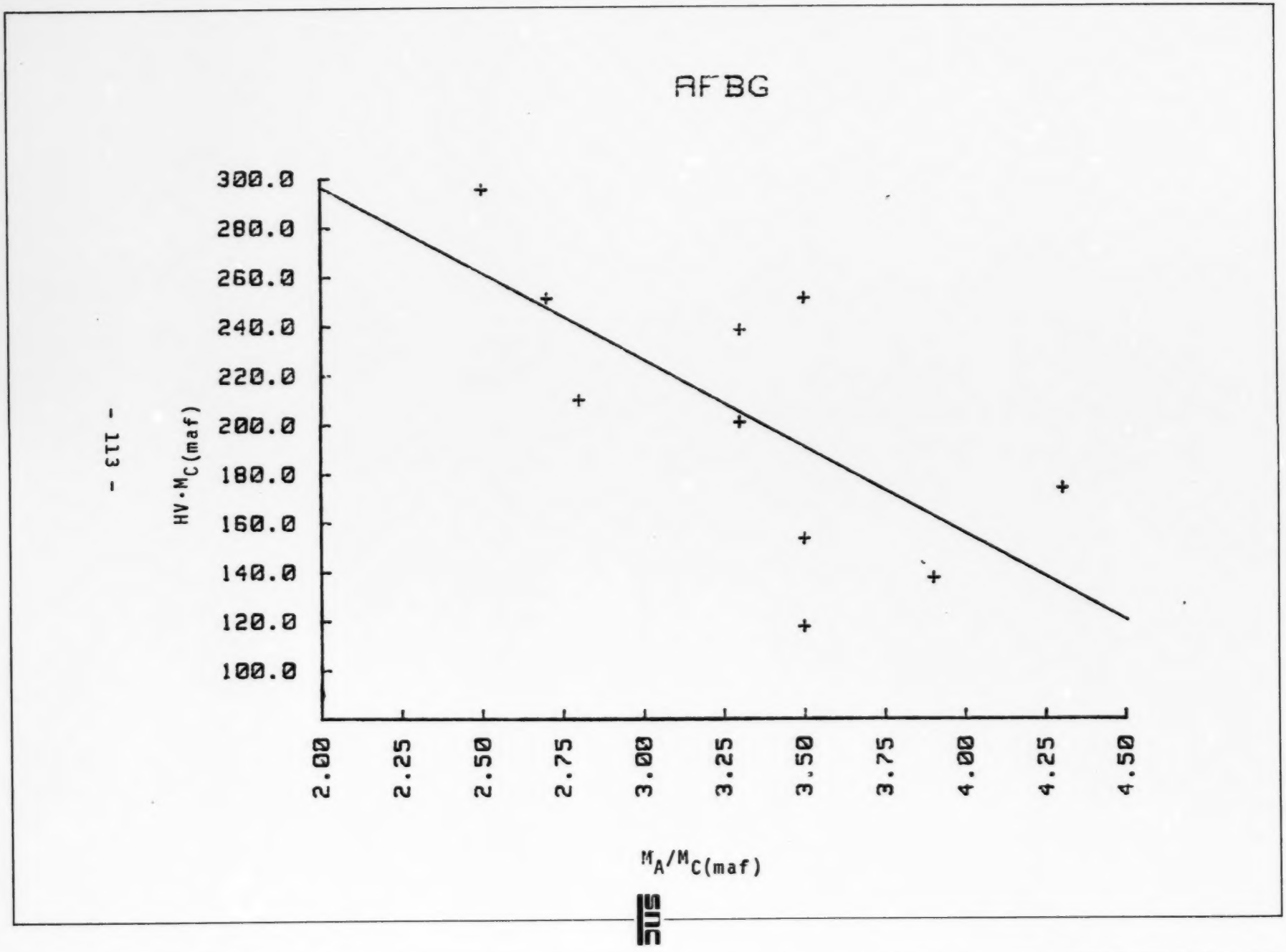

Sulfur Capture, % (Balance on Coal and Product Gas)
10
20
30
40
50
60
70
80
90
100
0.0
.1
.2
.3
.4
.5
.6
.7
.8
.9
1.0
1.1
1.2
1.3
1.4
$M_L/M_{C(maf)}$
AFBG


AFBG
Sulfur Capture, % (Balance on Solids Streams)
100
90
80
70
60
50
40
30
20
10
0
10
20
30
40
50
60
70
80
90
100
Sulfur Capture, % (Balance on Coal and Prdduct Gas)


AFBG
30.0
27.5
25.0
22.5
20.0
17.5
15.0
12.5
Carbon Conversion, %
2.00
2.25
2.50
2.75
3.00
3.25
3.50
3.75
4.00
4.25
4.50
$H_A/M_{C(maf)}$

AFBG
Gas Heating Value, MJ/m3
4.00
3.75
3.50
3.25
3.00
2.75
2.50
2.25
2.00
1.75
1.50
1.25
2.00
2.25
2.50
2.75
3.00
3.25
3.50
3.75
4.00
4.25
4.50
MA/MC(maf)

RFBG
Product Gas Yield, m³/kg(maf)
4.00
3.75
3.50
3.25
3.00
2.75
2.00
2.25
2.50
2.75
3.00
3.25
3.50
3.75
4.00
4.25
4.50
MA/MC(maf)
USS


AFBG
300.0
280.0
260.0
240.0
220.0
200.0
180.0
160.0
140.0
120.0
100.0
HV·MC(maf)
2.00
2.25
2.50
2.75
3.00
3.25
3.50
3.75
4.00
4.25
4.50
MA/MC(maf)

POOR

QUALITY

ORIGINAL